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# STUDY OF CHARACTERIZATION AND CONTROL OF AIR POLLUTANTS FROM A FLUIDIZED-BED COMBUSTION UNIT

# THE CARBON-BURNUP CELL

for

Division of Control Systems
Office of Air Programs
Environmental Protection Agency

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16. Abstracts The Carbon-Burnup Cell is a region of a fluidized bed boiler in which the fuel is the carbon bearing flyash carried out of the adjacent coal burning regions of the boiler. An effort was made to produce design criteria for an effective Carbon-Burnup Cell. The experimental work was conducted in two different test rigs. The more basic work, that which led to a statistically derived performance model, was done in a separate column which depends on an external source for its fuel supply. While this permitted the rate of fuel injection to be controlled it also meant that the flyash fuel was days or even weeks old. The second set of tests was performed in an actual fluidized-bed boiler, generating the flyash which is then immediately burned in its own internal flyash fired section, the prototype Carbon-Burnup Cell. The work in the boiler led to engineering insights on the design requirements for a commercial fluidized-bed boiler which included a Carbon-Burnup Cell. The operation of the boiler with a simulated Carbon-Burnup Cell also served as an effective demonstration tool for the process and design concepts. A coal-fired, fluidized bed boiler may be constructed which can operate at relatively high air rates ( 800 lbs/hr ft$^2$) and achieve 99% combustion efficiency. The desired level of combustion efficiency may be achieved by recycling carbon bearing fines collected from the flue gas of the coal burning regions to a region of the fluidized-bed boiler in which the bed temperature is in the range 1950-2050 F. The requirements for bed depth, firing rate and air rate for a Carbon-Burnup Cell may be determined using the performance model derived. Fuel costs are minimized; i.e., the system is optimized, when the boiler is designed and operated so that approximately 90% of the fuel value, fed as coal to the fluidized-bed boiler, is consumed in the primary cells and 10% in the Carbon-Burnup Cell(s). Optimizing fuel costs is equivalent to optimizing system costs. Calcium sulfate formed in the low temperature (i.e., coal-burning) regions of a fluidized-bed boiler by injection of fine limestone will not decompose in the high-temperature Carbon-Burnup Cell if the residual oxygen level is maintained at above 3.5%.

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| | |
|---|---|
| Air pollution control equipment | Efficiency |
| Boilers | Temperature |
| Fly ash | Expenses |
| Design criteria | Limestone |
| Coal | Calcium sulfates |
| Air | |
| Combustion | |

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# STUDY OF CHARACTERIZATION AND CONTROL OF AIR POLLUTANTS FROM A FLUIDIZED-BED COMBUSTION UNIT

## THE CARBON-BURNUP CELL

for

Division of Control Systems
Office of Air Programs
Environmental Protection Agency

under

CONTRACT CPA 70-10

E. B. Robison, R. D. Glenn, S. Ehrlich
J. W. Bishop and J. S. Gordon

FEBRUARY, 1972

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INCORPORATED

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D. Bruce Henschel of the Division of Process Control Engineering, Office of Air Programs, has been project officer for Pope, Evans and Robbins work since 1968.

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# APPENDICES

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D. Selected Carbon Burnup Results

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M. Why the Carbon-Burnup Cell?

# 1. SUMMARY

## 1.1 General

Pope, Evans and Robbins, in a continuation of a project sponsored by the Office of Air Programs of the Environmental Protection Agency (OAP), has monitored air pollutant emissions from the combustion of coal in a fluidized-bed boiler. This report describes the results of experiments and studies carried out between October 1969 and October 1970. In earlier work (1)* it was found that sulfur oxide emissions could be markedly reduced by injecting finely divided limestone into a coal-burning, fluidized bed operating at 1500°F to 1600°F and with about 3% residual oxygen** in the flue gas. However, with 3% residual oxygen in the flue gas, an economically unacceptable fraction of the input coal's fuel value would be lost as carbon in the flyash that is blown out of the furnace.

A comprehensive search for methods to reduce the loss of fuel, which ranged as high as 15%, led to the invention of the Carbon-Burnup Cell (2). The Carbon-Burnup Cell is simply a region of a fluidized bed boiler in which the fuel is the carbon bearing flyash carried out of the adjacent coal burning regions of the boiler. The Carbon-Burnup Cell is characterized by design features which permit operation at a relatively high temperature.

---

*Underscored figures in parenthesis refer to items in the list of references in Section 10 of this report.

**Values for "residual oxygen" are given in lieu of the more common "excess air" because the latter is ambiguous in discussing the fluidized-bed boiler. A 3% residual oxygen level is the equivalent of 15 to 17% excess air.

The term "cell", which has an interesting origin, is defined in Section 4. In this summary section it should be explained that a commercial fluidized-bed boiler will be composed of a combination of primary, burnup and, probably regenerator cells. These cells are distinct regions of the boiler in which specialized functions are carried out. In one design concept 75% of the fluidized-bed boiler's cross-sectional area is devoted to the coal-burning primary cells, about 15% to the flyash burning Carbon-Burnup Cells and 10% to regenerator cells in which calcium sulfate is decomposed. A simplified schematic diagram of such a system is shown in Figure 1.

OAP selected the Carbon-Burnup Cell for extensive study because realization of the air pollution control potential of the fluidized-bed boiler depended on the units commercial feasibility, which in turn depended on the efficient combustion of the fuel. The program described in this report consisted, for the most part, of an effort to produce design criteria for an effective Carbon-Burnup Cell.

The experimental work was conducted in two different test rigs. The more basic work, that which led to a statistically derived performance model, was done in a separate column which depends on an external source for its fuel supply. While this permitted the rate of fuel injection to be controlled it also meant that the flyash fuel was days or even weeks old. The second set of tests was performed in an actual fluidized-bed boiler, generating the flyash which is then immediately burned in its own internal flyash fired section, the prototype Carbon-Burnup Cell. The work in the boiler led to engineering insights on the design requirements for a commercial fluidized-bed boiler which included a

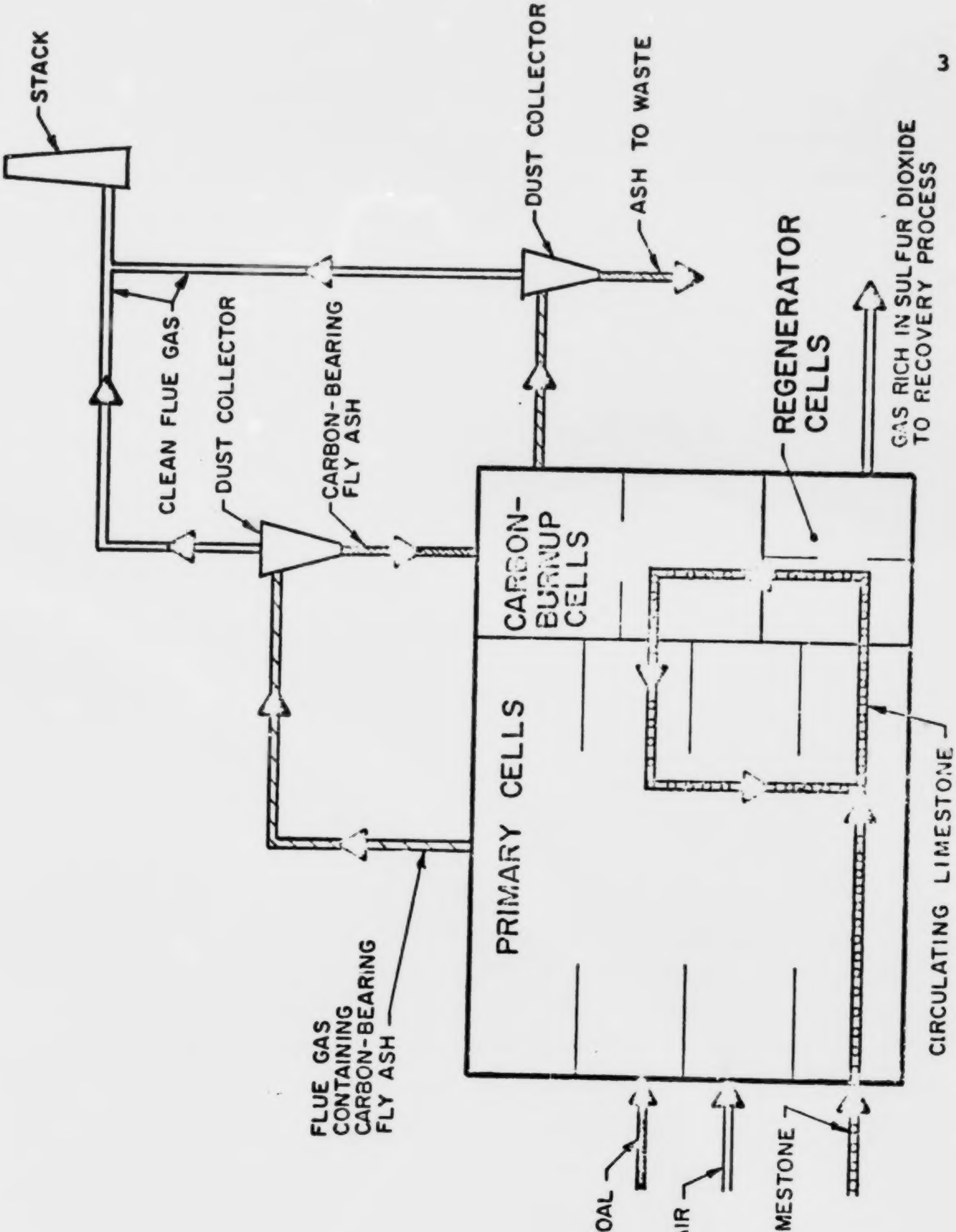

FIGURE I SIMPLIFIED SCHEMATIC OF MULTICELL FLUIDIZED-BED BOILER

Carbon-Burnup Cell. The operation of the boiler with a simulated Carbon-Burnup Cell also served as an effective demonstration tool for the process and design concepts.

The apparatus and the work are discussed below.

## 1.2 Tests in the Fluidized-Bed Column

The major portion of the effort to produce the design correlations for flyash combustion was carried out in the Fluidized-Bed Column (FBC). The FBC is an experimental fluidized-bed combustor which has a cross-sectional area of 0.86 square feet.

Carbon-containing flyash was burned in this device over a wide range of conditions, summarized as follows:

| Parameter | Range Studied |
|---|---|
| Bed temperature | $1750^{\circ}$F to $2140^{\circ}$F |
| Bed depth (static) | 10" to 22" |
| Air rate | 385 to 1000 lb/hr ft$^2$ |
| Fuel rate | 48 to 350 lb/hr ft$^2$ |
| Carbon content of the flyash fuel | 28% to 65% |
| Heat Removal Rate | 15% to 40% of heat release |

For any given test the specific values within these ranges were selected on the basis of a statistically-designed program of experiments. As discussed in Section 5.4.1, the carbon content of the flyash fuel was not controllable. The bed material in these tests was crushed, sintered coal-ash.

The model equations, based on an analysis of 38 FBC experiments are given below.

The equations are presented in a normalized form. The normalization assumes that the proper scaling factor is cross-sectional area. This has not been demonstrated and will not be until a large Carbon-Burnup Cell is constructed.

$$\begin{aligned}
&\text{Combustion efficiency, \%} = \\
&-13.78 \\
&+ 0.05193\ (\text{bed temperature, } {}^{\circ}\text{F}) \\
&+ 0.03973\ (\text{air rate, lb/hr ft}^{2}) \\
&+ 0.3831\ (\text{static bed depth, inches}) \\
&- 0.7514\ (\text{carbon feed rate, lb of carbon/hr ft}^{2}) \\
&- 0.1638\ (\text{inert feed rate, lb of inert/hr ft}^{2}) \\
&+ 0.0020\ (\text{carbon feed rate x inert feed rate,} \\
&\qquad \text{lb}^{2}\text{/hr}^{2}\text{ft}^{4})
\end{aligned} \tag{1}$$

The standard deviation is 2.4%.

This model has been tested using data obtained in runs not used in deriving the model. The results of these tests indicated that the model was not limited to simply reproducing itself but was probably a useful prediction tool.

It was further shown that by adjusting the control parameters, the combustion efficiency of mixtures of flyash and bituminous coal could also be predicted. Section 1.4 discusses another use for Equation 1.

The residual oxygen content of the flue gas could also be predicted, although somewhat less accurately than the combustion efficiency. This model equation is given on the following page.

A prediction tool for the $CO_2$ content of the gas also comes out of the statistical analysis. It was reassuring to note that the equation $C + O_2 = CO_2$ was reproduced by the model equations.

Residual Oxygen, % =
+22.91
-0.007353 (bed temperature, $^{o}$F)
+0.01118 (air rate, lb/hr ft$^2$)
-0.1390 (static bed depth, inches)
-0.1521 (carbon feed rate, lb carbon/hr ft$^2$)
-0.0151 (inert feed rate, lb of inert/hr ft$^2$)
+0.0002653 (carbon feed rate x inert feed rate, lb$^2$/hr$^2$ft$^4$) (2)

The standard deviation is 0.69%, based on the 38 tests in the FBC.

The body of the report contains a sample calculation which shows how one may use these two model equations in the design of a Carbon-Burnup Cell.

Mixtures of carbon-bearing flyash and partially sulfated fine lime were also fired in the FBC. These tests were conducted because it was possible to control $SO_2$ emissions by injecting finely ground limestone into the primary cells of a fluidized-bed boiler. The fine limestone, would calcine, be partially converted to calcium sulfate and then leave the primary cells as part of the flyash in flue gas. Most of this dust would be removed from the flue gas by some gas cleaning device, either a cyclone or electrostatic precipatator but never a wet scrubber. Since no practical method exists for separating the spent sorbent from the carbon particles or ash all of the dust collected would be injected into the Carbon-Burnup Cell.

The tests, in which the fuel to the FBC contained calcium sulfate, were intended to determine how, or rather if, the calcium sulfate content affected the carbon combustion process and in turn how exposure to the high temperatures in the FBC would affect the sulfate.

In the few tests conducted it was found that when the partially sulfated lime entered the FBC along with the carbon-bearing fly ash, its sulfur content could be released as $SO_2$ if the FBC operated at a temperature of approximately 2000°F with residual oxygen $\leq$ 2.0%. The highest $SO_2$ concentrations observed from the FBC during these tests, approximately 1/2%, were measured when the residual oxygen level was approximately 0.1%. On the other hand, when the residual oxygen level in the FBC was in the range 3.5 to 6.0%, the fine sulfate did not decompose at burnup cell temperatures. In fact, in one test at 6% residual oxygen the partially sulfated limestone appeared to still be reactive at a bed temperature of 1980°F. Two possible explanations of this result are provided in the body of the report. With high residual oxygen sulfur oxide emission from the FBC was as low as 350 ppm.

When no sorbent was present, the sulfur in the fly ash would burn with an efficiency equal to that of the carbon. This fact is of no special significance but it is interesting to note that when coal is burned at low temperatures in a fluidized-bed the carbon is more refractory than the sulfur.

The data gathered on nitric oxide emissions from a Carbon-Burnup Cell was correlated less accurately than other parameters, though the highest emissions were detected at the highest bed temperatures. A mean value of 539 ppm was measured for all fly ash combustion tests in which the bed temperature was above 1900°F. No efforts were made to reduce nitric oxide emissions which remains an area requiring research. However, even at the level of 539 ppm from the Carbon-Burnup Cell the fluidized-bed boiler, as a whole, compares favorably with conventional boilers firing coal.

Particulate emissions from the FBC were found to decrease with increasing bed temperature, possibly through agglomeration of ash matter to bed particles. The correlation produced by the statistical analysis of the independent variables used for Equation 1 and 2 is

Pounds of particulate emission per hour per $ft^2$ =

$$\begin{aligned} &+15.57 \\ &- 0.00664\ (\text{bed temperature, }^{\circ}F) \\ &- 0.00135\ (\text{air rate, lb/hr ft}^2) \\ &+ 0.03034\ (\text{carbon rate, lb/hr ft}^2) \\ &+ 0.00461\ (\text{inert rate, lb/hr ft}^2) \\ &+ 0.00034\ (\text{carbon rate x inert rate, lb}^2/\text{hr}^2\text{ft}^4) \end{aligned} \tag{3}$$

The residual standard deviation is 0.75 lb/hr $ft^2$

Since in some tests the fuel was approximately 65% ash, particulate emissions were remarkably low. The potential of retaining a large fraction of the ash in the bed of a Carbon-Burnup Cell should be explored further.

Hydrocarbons and carbon monoxide from the FBC were essentially nil when operating in the Carbon-Burnup Cell mode, i.e., high temperature and high excess air.

### 1.3 Tests in the Fluidized-Bed Module

Tests were also conducted in the Fluidized-Bed Module (FBM), an actual boiler, with a grate* of approximately 9 square feet.

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*The terms air distribution grid, grid, air distributor and grate are used interchangeably in this report.

A small (1.1 $ft^2$ cross-sectional area) test column was appended to the FBM in which the carbon bearing fly ash, produced by the FBM, could be fired. The purpose of these tests was to confirm that the results achieved in the FBC, under carefully controlled conditions, could be achieved, under the more realistic conditions of an actual boiler operation. With this addition the FBM had a grate area of slightly over 10 $ft^2$.

The experimental test column was designated the "CBC". In this report the term CBC is only used when referring to the specific item of experimental apparatus added to the FBM. The CBC only simulated a Carbon-Burnup Cell, as this is defined in Reference 2 and in Section 4 of this report.

While demonstration was an important part of the FBM/CBC design and test program it was also necessary to determine the problem areas in operating a fluidized-bed with two distinct temperature regions. A temperature of about 1500°F was to be maintained in the FBM, the CBC was to operate at 2000°F. Bed material could circulate between the two sections.

It was found that for the small CBC, an opening in the barrier between the two sections of just 2 square inches provided adequate bed material circulation and permitted the desired temperature difference to be achieved. An interchange rate between the two regions of 12,000 pounds of bed material per hour per square foot of opening was estimated from two transient heating tests described in the report.

The tests with the FBM were only a partial success; in most tests the combustion efficiency was less than 99%. However in the two final tests it was possible to demonstrate that the burnup results achieved in the FBC could be reproduced in an actual boiler. So it was shown that an overall combustion efficiency of 99% could be achieved. Some important insights, i.e. "know how", were obtained but the CBC had some significant flaws that made it difficult to continue a test for more than about eight hours. Subsequent to the work described in this report a second generation CBC was built after radical changes were made to the FBM. The new column, which retained the air distributor and fuel injector of the old, can operate at its design point continuously.

Some of the problems which made it impossible to perform the desired parametric studies before the program ended are outlined below.

A new coal feeder, required by the addition of the CBC to the FBM, performed very poorly until a number of minor alterations corrected the problems. The new feeder concept appears to be more practical for commercial use. Coal feeding to a fluidized-bed boiler is an area which requires further development effort.

Two fly ash feeders were tested in the CBC. One design, termed a "mushroom" feeder, performed well, giving a relatively even fuel distribution.

Two bed particle "knockouts" (one water-cooled, the other uncooled) were also tested in the CBC. These baffles both consisted of cylinders arranged in a horizontal triangular array placed just above the fluidized bed. Without a baffle screen, large quantities of bed material would be lost from the CBC; with either baffle screen, the quantity of bed material carried over was small. When the baffle screen was made up of water-cooled tubes, an initial heat transfer coefficient of approximately 35 Btu/ft$^2$ hr$^{\circ}$F was measured, where the referenced temperature is that of the fluidized bed. This was approximately twice the value that would be predicted from simple radiation and convection, indicating that an active heat transfer region exists above the dense phase of a high velocity (hence turbulent) fluidized bed.

When all of these changes had been made and operational problems overcome, the FBM achieved an overall combustion efficiency of approximately 99%.

When finely divided (-325 mesh) No. 1359 limestone was injected into the FBM, less sulfur was removed than in the tests described in Reference 1 (approximately 70% at Ca/S =2.5 compared to approximately 80% at Ca/S = 2.5 measured previously). The reasons for the less favorable performance are unknown but may have been the result of a different injector design. This problem was not pursued for reasons brought out later.

## 1.4 Use of Data Acquired in Both FBC and FBM

Data gathered in coal combustion tests during earlier work (1) at relatively low temperatures (1500-1750$^{\circ}$F) in both the FBC and the larger FBM allowed an equation for fuel energy loss, as high carbon fly ash, to be derived. Equation 4 shown below, links carbon combustion efficiency with the most significant control variable for combustion of a coal in a "low-temperature" fluidized bed.

Carbon Combustion Efficiency, % =

$$71.2 \times \left[\text{Residual Oxygen,}\right] \%^{\,0.145} \tag{4}$$

The standard deviation for this equation based on 28 data points is S = 4.1%.

Since it is the carbon combustion efficiency in the coal-burning regions (or primary cells) of the fluidized bed boiler which determines the quantity and composition of fuel fed to the Carbon-Burnup Cell, the combustion efficiency of the system as a whole may be described with Equations 1 and 4.

A parametric study was performed to determine the optimum split in duty between the primary cells and the Carbon-Burnup Cells. As shown in the body of the report, the lowest fuel costs are realized if the Carbon-Burnup Cell burns about 10% of the input fuel (i.e., if the carbon combustion efficiency in the primary cell is 90%). Referring to Equation 4, this is equivalent to saying that the primary cells should operate with about 3% residual oxygen in the flue gas.*

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*NOTE that the hydrogen and sulfur burnup efficiencies in the primary cell are essentially 100%. The "carbon combustion efficiency" as stated here includes burnup to both CO, $CH_4$ and $CO_2$, and is not congruent with "combustion efficiency" as defined by heating value (Btu/lb or Btu/hr).

## 2. CONCLUSIONS

Based on an analysis of the experimental work and the performance model derived under this program, the following conclusions were drawn:

a. A coal-fired, fluidized bed boiler may be constructed which can operate at relatively high air rates (≈ 800 lbs/hr ft$^2$, and achieve 99+% combustion efficiency.

b. The desired level of combustion efficiency may be achieved by recycling carbon bearing fines collected from the flue gas of the coal burning regions to a region of the fluidized-bed boiler in which the bed temperature is in the range 1950-2050°F. The requirements for bed depth, firing rate and air rate for a Carbon-Burnup Cell may be determined using the performance model derived in this study.

c. Fuel costs are minimized; i.e., the system is optimized, when the boiler is designed and operated so that approximately 90% of the fuel value, fed as coal to the fluidized-bed boiler, is consumed in the primary cells and 10% in the Carbon-Burnup Cell(s). As will be shown in the text optimizing fuel costs is equivalent to optimizing system costs.

d. Calcium sulfate formed in the low temperature (i.e., coal-burning) regions of a fluidized-bed boiler by injection of fine limestone will not decompose in the high-temperature Carbon-Burnup Cell if the residual oxygen level is maintained at above 3.5%.

The following are considered valid, significant but (unfortunately) tentative.

e. Decomposition of this calcium sulfate can be achieved by operation of the Carbon-Burnup Cell at low levels of residual oxygen. However, the $SO_2$ concentration achievable with fine lime in a high velocity apparatus appears to be relatively low, approximately 0.5%. No firm conclusions on the maximum $SO_2$ level may be drawn from this work since attempting to locate the maximum was outside the scope of this effort. However several patents were located for processes indicating how higher concentrations could be achieved.

f. While fine limestone injection does not lend itself to regeneration and reuse of the lime, (no method exists for separating fine lime from equally fine fly ash) the coal ash and lime may be reacted within the Carbon-Burnup Cell, to form uniform agglomerates which may be of commercial value.

No firm conclusions on the requirements of the agglomeration process or market potential for the product are available since these tasks were also outside the scope of this study.

## 3. RECOMMENDATIONS

In order to realize the air pollution control potential of the fluidized-bed boiler as rapidly as possible, the following further actions are recommended:

a. A set of cost and performance goals for fluidized-bed boilers should be established. This would include air pollution control goals on $SO_x$, $NO_x$, hydrocarbons, CO, halogens, particulates and plume opacity.

b. Pope, Evans and Robbins, together with one of the major boiler manufacturers and a public utility, should perform an engineering design for a large (600 to 1000 MW) coal-fired, fluidized-bed boiler which may meet these goals.

c. Based on the questions which arise about that particular design, an experimental program should be conducted to answer those questions.

d. If, based on the assessment of an actual design, OAP's goals will be met, a prototype boiler should be constructed and operated. The prototype should be large enough to produce from 10 to 40 MW.

e. Pending a decision to proceed with the plan outlined above, the following experimental work on the air pollution control aspects of the fluidized-bed boiler should be carried out:

1) Tests of the $SO_2$ Acceptor Process (3) should be conducted which will determine, for geographically matched limestones and coals, the required sorbent circulation and makeup rates.

2) Experiments which indicate the directions in which reduced levels of oxides of nitrogen may be achieved

should be initiated.

3) A definition of the particulate and plume opacity control requirements of a fluidized-bed boiler should be found in cooperation with a leading manufacturer of dust collection apparatus.

# 4. INTRODUCTION

## 4.1 Description of the Fluidized-Bed Boiler

### 4.1.1 Background: The Need for a New Technique for Burning Coal

Since 1962 Pope, Evans and Robbins has carried out design and experimental studies (under the sponsorship of the Office of Coal Research, U. S. Department of the Interior) aimed at reducing the cost of utilizing coal as a boiler fuel.

The initial studies were concerned, primarily, with improvements in the design of plants and boilers for industrial steam generation; i.e., systems which would be large enough to supply power and/or process heat to a factory, but too large to be used by a laundry or apartment house and too small to be used by a large electric utility.

Boilers in this size range were selected for development for a number of reasons which can be summarized by a single statement - a novel boiler could be commercially successful in this size range with less development expense. Success at this level could then lead to scaling both downward and upward.

Conventional methods of firing coal, as a fixed-bed on a stoker grate or as a suspension flame in a pulverized fuel burner, were not found to hold promise for major reductions in size and cost regardless of development effort.

It was found that the cost of an oil or gas-fired boiler was low when compared to coal-fired boilers with the same steam capacity primarily because of differences in furnace size. Oil and gas could be burned in a smaller furnace than could coal. For industrial sized boilers this difference meant that oil and gas-fired boilers could be assembled in a factory and shipped to the user's site by rail, while coal-fired boilers could not. The coal-fired boiler would have to be assembled piece-by-piece, often out-of-doors, at the user's site and by workers who sometimes required close supervision. A method of firing coal was therefore needed which would reduce the size of the furnace so that large coal-fired boilers could also be factory assembled.

An evaluation of alternatives led to the selection of the fluidized-bed boiler as the most promising method of achieving the goals of our sponsor.

The fluidized-bed boiler would be smaller because a higher volumetric heat release rate and higher heat transfer coefficients could be achieved in a fluidized-bed than in a conventional furnace; i.e., the coal would burn more intensely, in less space, and with faster heat removal.

Therefore, in 1965 a development effort was begun, again under the sponsorship of the Office of Coal Research (OCR). By 1966 a primitive experimental boiler had been built and operated thus proving that fluidized-bed combustion could be used to develop a commercial boiler. The results were sufficiently promising to warrant an actual fluidized-bed boiler being built. The construction of the FBM(See Figure 7 and Section 5.2) was planned and the scope of the study portion of the effort expanded to include conceptual designs for the much larger boilers required for central power stations. (See Reference 4 for a description of this work.) A major incentive here was the fact that the fluidized-bed boiler could burn poorer grades of coal than could the pulverized coal boiler. The other advantages of the fluidized-bed boiler which would apply to utility size units as well as the smaller industrial units were the high heat transfer coefficients, leading to reduced tubing requirements, and the high volumetric heat release rates leading to a smaller furnace.

Also in 1966 dolomite was mixed with coal being fed in an experiment with the FBC (See Figure 2 and Section 5.1). Now another reason that this new technique for burning coal was needed became evident-- pollution control.

Representatives of the National Center for Air Pollution Control (now the Office of Air Programs) observed some of these first crude experiments and in 1967 began to support the air pollution control aspects of the fluidized-bed boiler.

### 4.1.2 Defining a Fluidized-Bed Boiler

Almost from the beginning of fluidized-bed combustion (5, 6) an extensive patent literature has been developing on techniques for using this new tool to generate steam from fossil fuels. (Reference 7 is believed to be the earliest U. S. patent of this type.)

Only the effort by Albert Godel (8), out of over twenty, has been commercially successful and this exception is generally known as a fluidized-bed boiler although it differs substantially from the system being discussed in this report.

To avoid confusion with Godel's design Pope, Evans and Robbins referred to its design as "The Direct Contact Heat Transferring Fluidized-Bed Boiler" (9). This tedious nomenclature has been dropped because the definition of boiler "the part of a steam generator in which water is converted into steam" cannot be applied to Godel's most significant contribution: the only commercially successful fluidized-bed furnace.

So as to make very clear what is meant in this report by "fluidized-bed boiler" the following definition is presented.

A fluidized-bed boiler is defined as a system which meets each of the following criteria:

a) The system's primary function is the generation of steam. Therefore, the materials of construction, the mode of operation, the arrangement, auxiliary power requirements, etc., are consistent with existing practices and economics in the conventional boiler field.

b) The fuel is added to and burned within a turbulent bed which has been termed a fluidized bed.

c) A significant fraction of the heat released by the burning fuel is extracted by heat transfer surfaces in contact with the turbulent bed.

A fluidized bed, in turn, is defined as a mass of particulate solids held in suspension by an upward current of fluid such that the bed has zero angle of repose and exhibits certain other properties of a liquid. Among the liquid-like properties of a fluidized bed which are important to the boiler designer, are that the bed becomes well mixed, with sufficient agitation, and the bed material can be caused to flow about or out of the system without the aid of mechanical devices.

### 4.1.3 A Simplified Description of a Fluidized-Bed Boiler

A fluidized-bed boiler consists of an enclosure containing both boiler tubes and a layer, or bed, of granular solids. The bottom of the enclosure is perforated so that air may enter the enclosure to cause the solids to become suspended, and to react with coal which will be added to the bed. Such a system is shown schematically in Figure 2. In addition to the

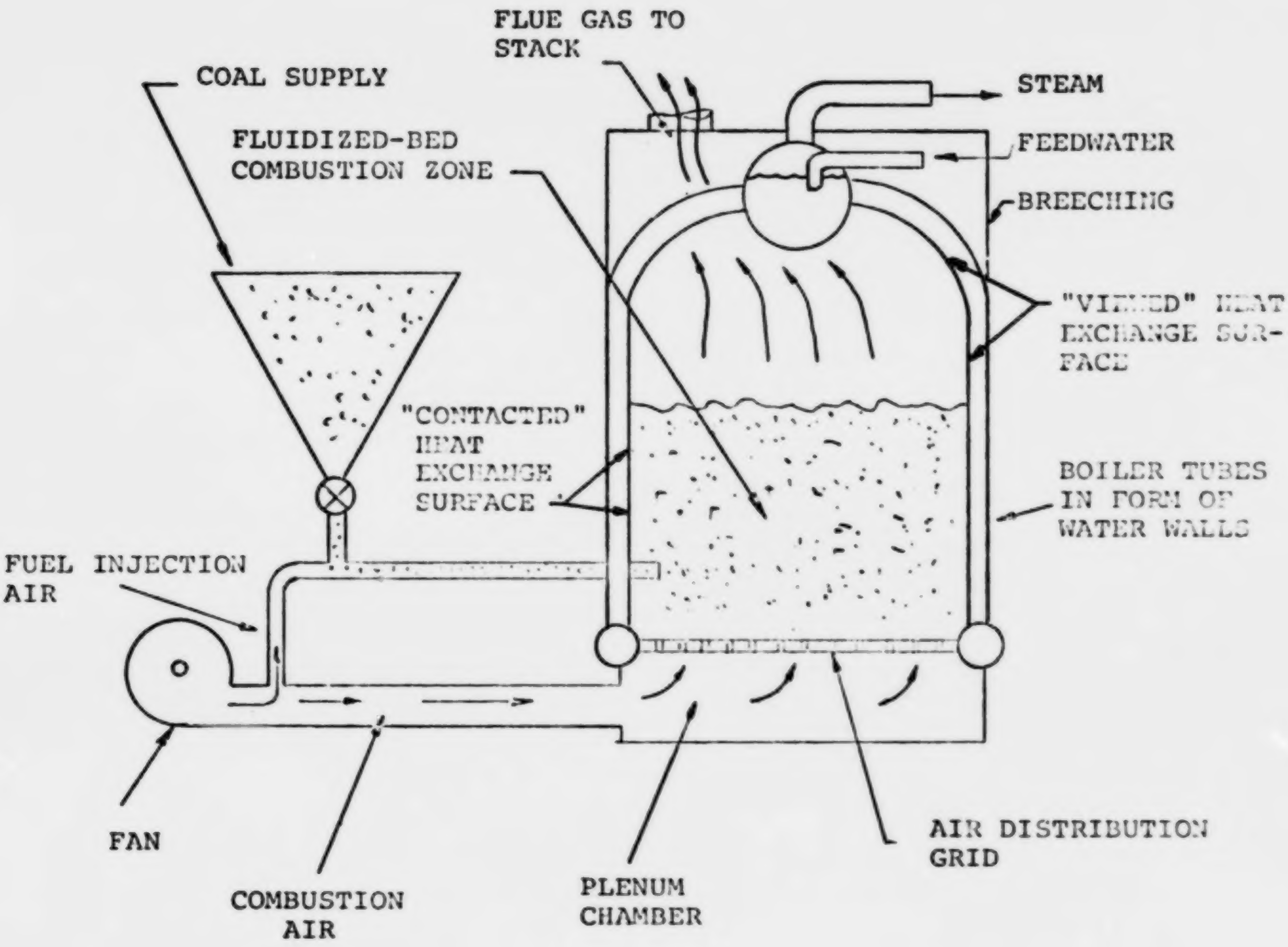
FLUE GAS TO STACK
COAL SUPPLY
STEAM
FLUIDIZED-BED COMBUSTION ZONE
FEEDWATER
BREECHING
"VIEWED" HEAT EXCHANGE SURFACE
"CONTACTED" HEAT EXCHANGE SURFACE
BOILER TUBES IN FORM OF WATER WALLS
FUEL INJECTION AIR
FAN
COMBUSTION AIR
PLENUM CHAMBER
AIR DISTRIBUTION GRID

FIGURE 2. SCHEMATIC OF FLUIDIZED-BED BOILER

components shown in the sketch, there would be convective heat exchangers and a system to remove particles carried out of the combustion space with the flue gas.

When the temperature of the bed is raised by an auxiliary means, such as a gas burner, to above about 800°F, bituminous coal added to the bed will ignite. After ignition the temperature will rise until the system achieves a thermal equilibrium; i.e., the energy added to the bed by the burning fuel precisely equals the energy extracted by the boiler tubes contacting and viewing the bed and by the gases and dusts leaving the bed. Because the coal burns very rapidly the fluidized-bed consists, almost entirely, of inert particles with a small quantity of coal. Typically a sample removed from the bed would be from 0.1% to 1.0% carbon by weight.

The equilibrium temperature may be as low as 1200°F or as high as 2500°F, although a narrower range, 1450°F to 2050°F, is of practical interest. Below about 1450°F it is difficult to consume the carbon monoxide or to prevent refractory hydrocarbons from appearing in the flue gas. Above about 2050°F the bed particles may agglomerate into large pieces and the bed collapse.

Assume that the simplified fluidized-bed boiler shown in Figure 1 is operating with a bed temperature of 1500°F. To increase the bed temperature additional coal and air might be added but it is more effective to reduce the depth of the bed by removing some of the bed material. This may not seem obvious but it should be recalled that the temperature at which the bed operates is that temperature at which the heat losses (which are proportional to the bed's temperature) precisely equal the heat input by the burning fuel. To increase the bed temperature one may either add more heat or remove less heat.

If the bed depth is reduced less of the boiler tubes are contacted by the bed and less heat is removed by these tubes. Since the heat input has remained constant the only way in which the heat losses can equal the heat input is for the bed temperature to rise. Then most of the extra heat available can be carried off by the hotter gases and dust.

The amount of heat transfer surface in the furnace has not been changed; but a tube which is contacted by the bed removes more heat than a similar tube which only views the bed.

In a commercial fluidized-bed boiler the bed temperature will not be changed and the bed depth will remain constant but the preceding example, hopefully, illustrates how bed depth and heat transfer surface are related and how they are used in design.

The design depth of the fluidized-bed may vary over a wide range - from a few inches to many feet. However, practical systems must operate in the range of about 12 to 47 inches. Below about 12 inches the combustion efficiency degrades and will become unsatisfactory; above about 48 inches the power required by the blowers becomes excessive since the power consumption will be proportional to the depth of the bed.*

Having set the operating temperature and bed depth in the fluidized-bed boiler, the energy release rate still remains unspecified. Rather than specify an energy, or heat, release rate, it is more appropriate to specify an easily measurable equivalent, the mass flow rate of combustion air, in pounds of air per hour per square foot of grate area. In conceptual designs air rates between 100 lb/hr $ft^2$ to over 2,000 lb/hr $ft^2$ have been considered. Practical systems will operate with between 400 and 1200 lb/hr $ft^2$.

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*Fluidized-bed boilers might be incorporated into combined cycles in which the hot flue gas, at high pressure, is passed through a turbine which is used to compress the combustion air as well as drive an alternator. In this system the economics are not so sensitive to bed depth. Supercharged boilers might logically evolve from the work described here.

Specifying the air rate is the equivalent of specifying the heat release rate because in a fluidized-bed boiler the rate of fuel injection is almost precisely set by the air rate. This statement assumes that a practical excess air level of about 10 to 20% is used and that the fuel is known. For every oil or coal there is a proper, albeit different, fuel-to-air ratio. The point of specifying air rate as the primary factor is brought out in the following paragraphs.

From a specification of the air rate, the bed depth and bed temperature plus the energy release per unit air, it is possible to compute the apparent plan and volumetric heat release rates, the superficial gas velocity and residence time and other parameters of interest. For example an air rate of 1000 lb/hr $ft^2$ is usually the equivalent of a superficial gas velocity of 12 to 15 fps, a plan heat release rate of about $10^6$ Btu/hr $ft^2$ and, in a five foot high furnace, a volumetric heat release rate of 200,000 Btu/hr $ft^2$.

The air rate selected determines the size and density distribution of the particles which will make up the bed. Particles above a certain maximum size and density will sink to the air distributor. Particles below a certain minimum size and density will be entrained in the gases leaving the bed and will be carried out of the system. The particles which make up the bed may be any noncombustible, granular solid which is sufficiently tough to retain its shape and size in the bed over an extended period.

In some coals a portion of the ash, (small stones, shale particles, etc.) is coarse enough and strong enough to remain in the bed and the coal can be burned in a fluidized-bed composed of its own ash. In fact, for this type of coal bed material must be continuously removed or the bed depth will increase. For other coals the ash may be too fine to be retained in a fluidized-bed, even if the coal itself is quite coarse, unless the ash is deliberately sintered. For the coals which cannot maintain their own bed, material must be added with the coal to make up for particles of the starter bed which are lost to the system through attrition. This material, besides meeting the properties listed above, must also be inexpensive. In many areas limestone may be the material of choice because of its cost and relatively low density* which permits a relatively deep bed to be used with a moderate pressure drop through the bed. Limestone possesses other properties of interest to the designer of a fluidized-bed boiler and these are discussed in the following paragraphs.

## 4.2 Air Pollution Control Potential of Fluidized-Bed Boilers

### 4.2.1 Perspective

Fluidized-bed combustion of coal is, in itself, not a remedy for air pollution; a fluidized-bed boiler is not a pollution control device. However, certain properties of fluidized-bed combustion and of a properly designed fluidized-bed boiler can be exploited to produce a steam supply which is

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*The limestone is dense but it soon becomes lime when it is fed into a fluidized-bed boiler. This report uses the term limestone for lime, or calcined limestone, when precise definition is not required.

"clean" at less cost than available alternatives.

What follows is intended to provide an overview of the air pollution control potential of the fluidized-bed boiler in the view of the authors of this report.

A report (to be published for EPA in 1972) by Westinghouse Research Laboratories entitled "Evaluation of the Fluidized-Bed Combustion Process" contains a summary of all of the available data and should also be read by anyone interested in the state-of-the-art.

### 4.2.2 Discussion

Air pollutants are what the laws define as air pollutants. For steam generators (boilers), air pollutants proposed and promulgated in 1971 are sulfur oxides, nitrogen oxides, the mass of the particulates and the plume opacity; i.e., the "look" of materials leaving the stack.

It is likely that before too many years have elapsed, rules will be established limiting the amount of arsenic, cadmium, mercury and other trace pollutants (e.g., beryllium) which can be emitted and just as for the automobile, carbon monoxide and hydrocarbons will be limited by law. For hydrogen chloride and chlorine, which have been mentioned, limits may be established some day.

A fluidized-bed boiler may be able to meet most of these regulations because of certain inherent characteristics of the fluidized bed combustion process.

Carbon monoxide and hydrocarbons are always produced while coal is being burned in a limited space. The FBC program was not undertaken primarily to control CO and hydrocarbons, since conventional utility boilers also have low emissions of these pollutants due to their large volume. They enter the atmosphere when the burning process is not complete. By careful design, partial combustion products can be consumed within the fluidized-bed and in the freeboard. And although no work appears to have been done on methods of reducing CO and $C_xH_y$ (beyond decreasing the coal:air ratio), emissions on the order of 100 ppm for each can be anticipated for fluidized-bed boilers. Additional research may provide methods for a further reduction in emissions. A low-cost combustion catalyst approach may be economical in a fluidized-bed boiler, but is less likely to be practical for conventional boilers.

By the use of a tough limestone as the bed material, it is possible to absorb virtually all of the sulfur released by the coal. To achieve very low levels of $SO_2$ emission requires that a very large excess of active lime be present in the bed and freeboard at all times. Economics, in turn, require that the limestone be kept active by continuously stripping off much of the sulfur so as to regenerate the limestone. The regenerator off-gas can then be processed to recover sulfur or sulfur products or be scrubbed, using a portion of the lime produced by the boiler. The property of a fluidized-bed boiler which makes effective in situ sulfur control possible is the relatively low temperature of the medium in which the combustion occurs. The bed may be kept cool by immersing heat exchange surfaces in the combustion bed or, alternatively, by circulating the sorbent bed through a separate heat exchanger. In addition, the refluxing of sorbent particles in the freeboard provides a degree of concurrency to an otherwise well stirred system. Additional development to provide countercurrency may make it possible to limit sulfur oxides emissions to 50 ppm. If a wet scrubber were added to control particulate emissions, the sulfur oxide emissions might be so low that the boiler could not longer, reasonably, be considered a source of sulfur oxides.

Nitrogen oxides are produced in a fluidized-bed boiler through the fixation of atmospheric nitrogen and the oxidation of nitrogenous compounds in the coal.* The nitrogen in coal is converted principally to gaseous $N_2$. Most coal-fired boilers produce less than 1000 ppm $NO_x$.

Because of the relatively low temperature of the combustion medium (about 1500°F compared to 3000°F in the conventional flame), it had been anticipated from equilibrium data on the fixation of atmospheric nitrogen that $NO_x$ from a fluidized-bed boiler would be very low compared to conventional boilers. However, this was not found to be the case, apparently due to oxidation of nitrogenous coal compounds. Measured values of $NC_x$ from most fluidized-bed combustors have usually been only moderately lower than emissions from conventional coal-fired boilers.

Little work has been done on methods of reducing $NO_x$ emissions from a fluidized-bed boiler, but it is hoped that research to be carried out in the future will lead to means for significantly reducing $NO_x$ emissions.**

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*Argonne National Laboratories, in an experiment utilizing an "artificial air" where argon was substituted for nitrogen, demonstrated that nitrogen oxides can arise from the nitrogenous compounds in the coal (10).

**Pope, Evans and Robbins showed, in a recent test series, that more NO resulted from combustion of 3/4" coal than from combustion of 1/4" coal. If only fine coal is burned, the carbon inventory in the bed is minimized.

For the control of particulate emissions, the fluidized-bed boiler possesses certain inherent advantages over most conventional boilers. Since the coal fed to the fluidized-bed boiler is crushed and not finely ground, some of the larger pieces of ash termed non inherent will not be entrained in the gas at all. Further, most of the particles which are entrained may be removed by a relatively inexpensive mechanical dust collector. Such a collector, while inefficient on a pulverized-fuel furnace, is efficient on the ash from a fluidized-bed boiler.

Inherent ash, on the other hand, may be very fine even if the coal is very coarse* and will be entrained by the gas. Some of this fine material will pass through a mechanical dust collector and could be removed by an electrostatic precipitator. This final collection is potentially difficult when sulfur oxides** in the gas are low. However, in a fluidized-bed boiler in which a Carbon-Burnup Cell is used, the high carbon content of the fly ash from the primary cells should act as a natural resistivity conditioning agent. This should allow efficient dust removal at conventionally low temperatures from the gas leaving the primary cells. The dust fed to the Carbon-Burnup Cell could be caused to sinter into larger pieces and that which is emitted could be removed by a bag filter without a significant cost penalty (less than 1% of the plant costs).

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*About 10 pounds of coal was taken from a pile.. The largest pieces were over 1". The coal was put in an ashing furnace and heated carefully so that the ash would not melt. The ash was then carefully screened and weighed. About 1/4 of the ash was larger than 1/2" with one piece 1", about 15% was between 1/20" and 1/2". The rest, 60%, was less than 1/20" and most was as fine as face powder.

**It has been shown that, although the electrical resistivity of most of the ash matter (silica, alumina, and ferric oxide) is high at conventional collection temperatures, sulfur oxides which would normally be in the gas reduce the resistivity to the point where the electrostatic precipitator can function efficiently.

A second standard for particulates is concerned with the opacity of the plumes issuing from the power plant's stacks. "Smoke" control laws were the first promulgated* and may be the most difficult to meet in the long run. Standards now proposed would require that the emission be "invisible". While the opacity of the plume is related to the weight of particulates, it is not congruent with weight. A relatively opaque plume may result, for example, when firing conditions result in incomplete combustion causing micron-sized carbon particles to form from cracked hydrocarbons. Gas turbines produce smoke through this mechanism. This will not occur in a properly designed fluidized-bed boiler.

A second source of light-scattering particles (causing the plume to be visible) is sulfur trioxide droplets formed as the dew point is reached in the gas stream. When the bed of the fluidized-bed boiler contains limestone, the quantity of sulfur trioxide is vanishingly small.

A third source of small particles is the inorganic fumes and smokes which result when mineral matter is volatilized during combustion. Because of the low temperature

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*One of the earliest smoke control laws provided for beheading of violators. This penalty has not yet been suggested as the solution to our current pollution problems.

(about 1500°F) of the combustion medium in a fluidized-bed boiler, less mineral matter will volatilize than in a conventional boiler flame. As much as 99% of certain normally volatile species can be retained in a fluidized-bed boiler operating at low temperatures.

Mercury has been declared a hazardous pollutant. Standards for $CL_2$ - HCL, among others, are being contemplated.

The pollutants over which there is no current (1971) concern but which may soon be considered are heavy metals and radioactive isotopes. There is some evidence that arsenic and beryllium would be retained in a bed of lime, but cadmium and mercury would not be retained. Radioactive species should be lower from the fluidized-bed boiler than from a pulverized fuel unit but little thought has been given to this. Chlorides have not been measured but a plausible set of chemical reactions and some intuition suggests that they may not appear as gases from a fluidized-bed of lime.

From the above it may be seen that the air pollution control potential of the fluidized-bed boiler is quite promising. This potential follows from certain inherent characteristics: the low temperature of the bed, the low temperature of the combustion, the high carbon content of the fly ash in the main flue gas stream, the stability of the combustion, the ability of lime to react with volatile inorganic materials to form stable solids or liquids.

The first generation of fluidized-bed boilers may not achieve all of the goals set in the preceding paragraphs. However, by the careful and ingenious application of tools made available by this new form of steam raising it may be possible to evolve designs that will provide clean and economical power.

## 4.3 Pope, Evans and Robbins' Prior Work

### 4.3.1 General

Pope, Evans and Robbins, under contract with the Office of Air Programs (OAP) of the Environmental Protection Agency (formerly the National Air Pollution Control Administration, Department of Health, Education and Welfare), has characterized emissions from a fluidized-bed boiler developed for the Office of Coal Research, Department of the Interior. Two reports have been prepared describing this work (References 1 and 4).

The pollution control aspects of a fluidized-bed boiler had been considered as early as 1965 when it was discovered that if coal distribution was uniform, smokeless combustion was achieved at very low excess air levels. As noted before, in 1966 a test was conducted in which dolomite was mixed with the coal and the sulfur oxide control potential of the process was demonstrated. A literature and patent search revealed that the use of limestone in a fluidized-bed combustor to control sulfur oxides from a coal-fired boiler was novel, though a patent had been granted earlier on the use of limestone in an analogous process.*

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*The use of limestone in a coal-fired, fluidized-bed boiler should not infringe this existing patent which was narrowly drawn.

### 4.3.2 Once-Through Limestone Injection

Based on this early work, the initial studies conducted for OAP investigated the use of relatively coarse limestone and dolomite (crushed to -7+14 mesh) on a once-through basis. Once-through is the term used to describe a non-cyclic or non-regenerative process. The limestone is added to the bed and allowed to blow away or overflow and is not reused. Although in a fluidized-bed boiler the dust which blows out of the primary cells is returned to the Carbon-Burnup Cell, it would still be a once-through process if the sorbent is not regenerated and reused.

The bed in these tests was primarily coal ash and the air rate was chosen to yield a nominal superficial velocity of 10 to 15 feet per second. The coarse limestone particles fed to the bed would not elutriate at these velocities until they were reduced in size by attrition. It was determined that coarse stones would not retain nonregeneratively more than about 30% of the sulfur released by the coal at reasonable rates of sorbent addition.

Each test in this series was run long enough to assure that a steady state value of the sulfur oxide emission had been achieved, no more than two or three hours and often much less. Depending on the coal and stone used, the physical composition of the bed may or may not have been at steady state.

If a tough limestone and a well washed coal were used the steady state bed, almost all limestone, would take over a hundred hours to establish. With the dolomite used in these tests, which decrepitated easily, the steady state bed, which contained little dolomites was achieved in only a few minutes.

One unexpected result of these first tests was that the capture rate decreased with increasing bed temperature. Bench-scale studies by others in which $SO_2$ was removed from a gas stream by a fixed bed of lime had indicated more rapid capture at 1800°F than at 1500°F. It had been assumed that the injection process into a fluidized-bed boiler was kinetically limited, yet the capture of sulfur was far better at 1600°F than at 1800°F.

In order to enhance the process kinetics, the stone was finely ground to pass through a 325 mesh screen. Although such a fine powder is rapidly elutriated from the bed at the air rates used here, the efficiency of sulfur capture was much higher with the finely ground sorbent. It was determined that about an 80% reduction in $SO_2$ emissions could be achieved by injecting a finely-ground limestone (identified by OAP as No. 1359) at Ca/S = 2.5. Raw stone and hydrate performed equally well, while stone calcined by the supplier performed poorly.

### 4.3.3 Regenerative Limestone Process

Towards the end of the pollution-characterization effort (May 1969), we found that beds composed almost entirely of the No. 1359 limestone could be made to release the sulfur that had been captured during the first few hours of a test in just a few minutes. This was done by increasing the coal-feed rate so as to increase the bed temperature and decrease the oxygen content of the flue gas. A regenerative cycle was thereby devised and a patent application prepared for the "$SO_2$ Acceptor Process" (3). This process was particularly attractive for application to a fluidized-bed boiler, while the previous state-of-the-art for decomposing calcium sulfate either damaged the limestone so that it could not be reused, were costly to apply, or exchanged pollution as carbon monoxide and hydrocarbons for pollution as sulfur oxides. This discovery was also felt to provide an explanation for the anomalous temperature behavior of the lime-sulfur oxide reactions noted earlier. In this hypothesis, sulfur retention in a fluidized-bed combustor was seen as an easily reversible process, sensitive to temperature, oxygen partial pressure and the presence of reducing gases.

### 4.3.4 Pollutants Other than $SO_2$

Pollutants other than sulfur dioxide were characterized --nitrogen oxides, hydrocarbons and particulates.

Oxides of nitrogen were found to increase as the amount of oxygen remaining in the flue gas increased. Experiments in which some air was diverted from the base of the bed to a port above the bed resulted in a marked reduction in NO. This might indicate that increased reducing gases in the bed were decomposing the NO. Nitrogen oxides were not materially and reproducibly affected by the injection of fine limestone.*

No levels of less than 400 ppm were found in this first test program; 275 ppm was typical.

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*Other workers have shown a reduction in NO with limestone injection (10). An increase in NO with limestone injection has also been predicted on the basis of experiments at Esso Research and Engineering (11). These apparent contradictions indicate that the statement - fluidized-bed combustion of coal is not an adequate description of the process. Each experimenter's reactor must behave differently. Since no attempt has been made to quantify these differences (for example, bubble frequency and diameter), an understanding of how the various results relate to boiler design parameters is not available.

Hydrocarbons and, presumably, carbon monoxide were found to be sensitive to the quality of the fuel distribution and to the coal:air ratio. With between 3% and 4% oxygen remaining in the flue gas, the hydrocarbons were reduced to below 100 ppm. Hydrocarbons were not affected by fine limestone injection.

The major fraction of the mineral matter in the coal normally appears as fly ash. The larger particles of non-inherent ash remain in the bed. The starter bed material adds some particulate matter to flue gas, and when fine limestone is injected, essentially all of this appears in the fly ash stream.

At the high dust loadings, when fine limestone was injected, a low-pressure-drop mechanical collector removed about 90 wt. % of the fly ash. Without limestone the collector was 90% efficient. Therefore, with or without limestone injection, the dust emission was on the order of 1.5 pounds of particulate per million BTU of fuel input. More efficient dust collectors are anticipated for a commercial application.

The particles not removed by the collector were all smaller than 20 microns. It is these particles which would have to be removed by an electrostatic precipitator, bag collector, or wet scrubber.

### 4.3.5 Recommendations Based on Prior Work

The previous phase of the program was ended with recommendations for further study of the most promising leads. Sixteen of these recommendations were listed in Reference 1. One of these sixteen recommendations was the characterization of the carbon burnup cell, a region of a fluidized-bed boiler in which carbon-bearing fly ash is the primary fuel, as a means of promoting high combustion efficiency. This suggestion formed the basis of the work reported here. It was carried out during the period October 1969 through October 1970.

## 4.4 Prior Work Leading to the Burnup Cell Concept

Economics demand that commercially viable fluidized-bed boilers will have superficial gas velocities (primary cell) in the 5-10 ft/sec range; otherwise the size and cost of the unit is excessive. Economics also demand that commercially viable fluidized-bed boilers will burn essentially all the coal delivered to the boiler; i.e., fine coal as well as 3/4" coal. One of the economically attractive features of the fluidized-bed boiler is low coal preparation cost, since the coal need not be finely ground.

Bench scale studies by others have indicated that if freeboards are excessively large and superficial gas velocities are unrealistically low (e.g., 2 ft/sec) there is less need for a carbon burnup cell since the primary cell effluent carbon content is low by virtue of long residence time in the primary cell.

Authorities are generally agreed that if only large coal lumps (e.g., 3/4") were fed to a fluidized-bed boiler, the need for a carbon burnup cell would be less, by virtue of long coal residence time in the bed. Authorities are generally agreed that there is no realistic way to feed only 3/4" size coal lumps to the fluidized-bed boiler, since all of the heating value in a 3/4" top size lot of coal must be recovered, and metering screw feeders cause additional attrition.

Particles of coal, like the other particles which make up the fluidized bed, will appear in the freeboard either because they are small and are elutriated from the bed, or because they are thrown out of the bed by violent agitation. For further discussion of this topic, see Appendix L.

Based on this discussion, it is clear that if particles of coal, fed to the bed, have not been totally consumed before they come to the top of the bed, some coal particles will appear in the freeboard. It is also clear that the freeboard is an integral part of a fluidized-bed system. It is also clear from this discussion that unless the coal particles return to the bed or are totally consumed in the freeboard, unburned carbon values will be chilled by contacting cooled surfaces and will leave the furnace. An examination, over a period of several years, of possible ways to utilize these carbon values, led to the development of the two-stage combustion system incorporating a carbon-burnup cell.

## 4.5 Specific Objectives of This Work

As noted in Section 1, SUMMARY, the primary objective of this work was to demonstrate the Carbon-Burnup Cell concept and to develop a design correlation which would aid in the design of an actual Carbon-Burnup Cell.

A second objective was the derivation of a technique for optimizing the performance of a coal-fired, fluidized-bed boiler. Basically this requires a method to select the division in duty between the primary coal-burning cells and the flyash-burning Carbon-Burnup Cells so as to maximimize the boilers' thermal efficiency.

Some specific factors to be considered included;

injection of limestone into a burnup cell; the effect of the high temperature of the cell on partially sulfated limestone present in the flyash, especially decomposition; and the effect of fine limestone by-products on combustion efficiency.

It was not found practical to study the effect of horizontal heat transfer tubes within the bed of the high temperature burnup cell. The experimental apparatus was too small to permit a realistic tube array to be installed without absorbing an excessive fraction of the combustion energy. Two movable vertical water-cooled tubes were used to vary bed temperature so that some insights were achieved on the effects of heat removal. A larger apparatus, in which wall losses were relatively smaller, or a small apparatus with heated walls might each be useful for studies with horizontal tube bundles.

Design criteria for a fluidized-bed Carbon-Burnup Cell were to be provided. The effects of the burnup cell on the design and economics of Pope, Evans and Robbins' Multicell Fluidized-Bed Boiler concept were to be identified. The study was to be conducted in two parts. First, a statistical experimental program on the FBC. Second, a program of confirmation on the FBM. The scope of work prepared by OAP outlined the variables to be considered and these, as well as others, are discussed in some detail later in the report. (See Section 6.2.2).

# 5. APPARATUS AND MATERIALS

## 5.1 Pilot Scale Combustor, FBC

The majority of the tests were conducted in a pilot scale combustor, designated Fluidized-Bed Column (FBC). The FBC consists of a water-walled combustion chamber with plan dimensions of 12" x 16". A flat air distributor forms the bottom of the chamber and a vertical 12" diameter duct forms the outlet. Front and side elevations are shown in Figures 3 and 4.

In operation, air at ambient temperature flows into the plenum below the air distributor, up through the holes in the distributor and into the combustion chamber where it fluidizes the bed material and provides the oxygen for combustion. The bed material for the present tests consisted of sintered coal ash crushed and screened to the desired size. Fuel, coal and/or carbon-containing fly ash, is injected through a port at the base of the bed.

The air distributor contains a matrix of stainless steel bubble caps (buttons) mounted in a mild steel plate. A typical button is shown in Figure 5.

The bed temperature is monitored with a number of thermocouples spaced vertically in the combustor. Kaowool seals were provided between flanges to prevent flue gas leakage out of the system. Specifications for the FBC are presented in Appendix A.

The fuel feed system is capable of delivering approximately 250 pounds of coal each hour which would be sufficient for a firing rate of $3.5 \times 10^6$ Btu/hr. The air feed system is capable of delivering oxygen sufficient for

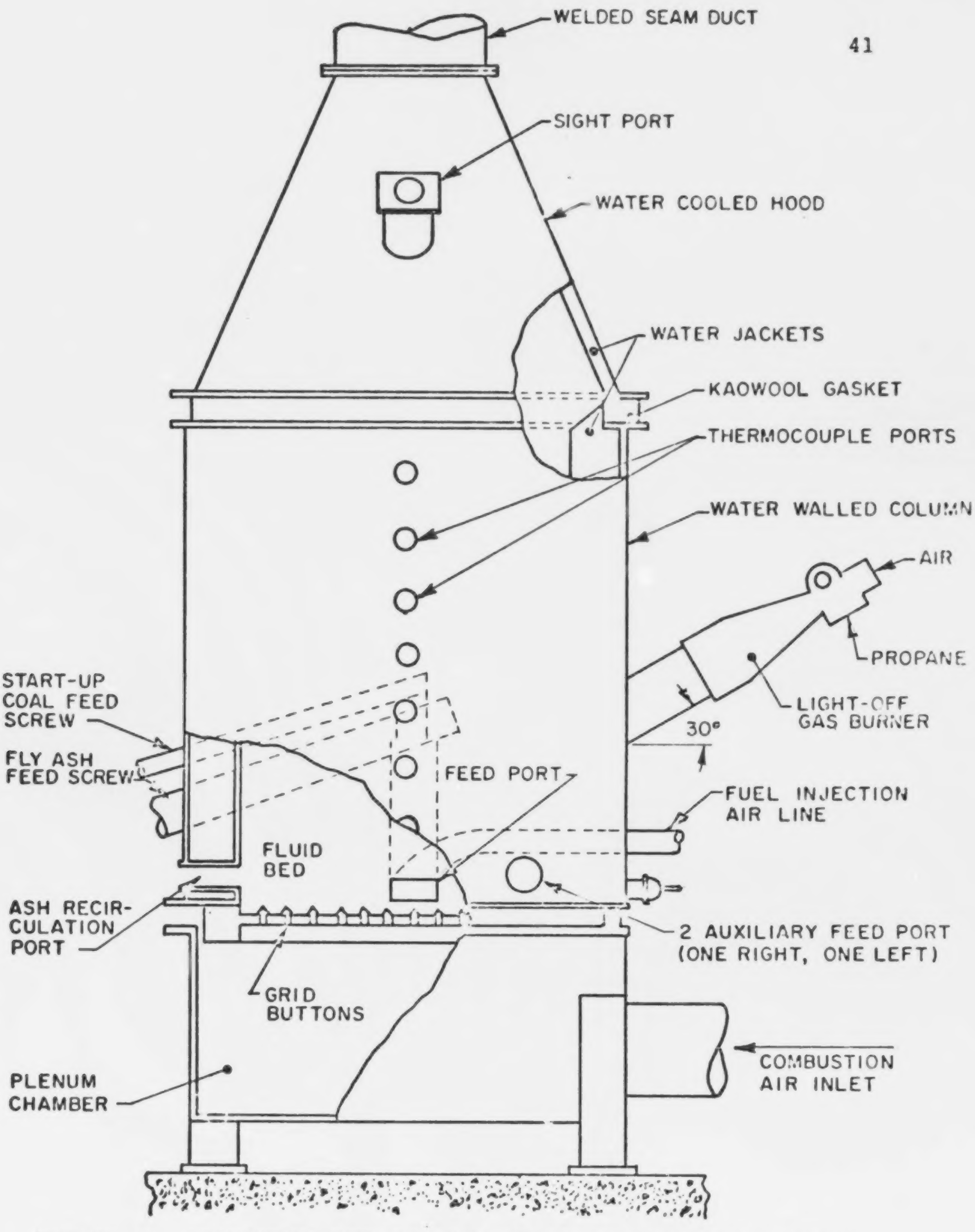


FIGURE 3. FLUIDIZED-BED COLUMN (FBC) CONSTRUCTION DETAIL FRONT VIEW

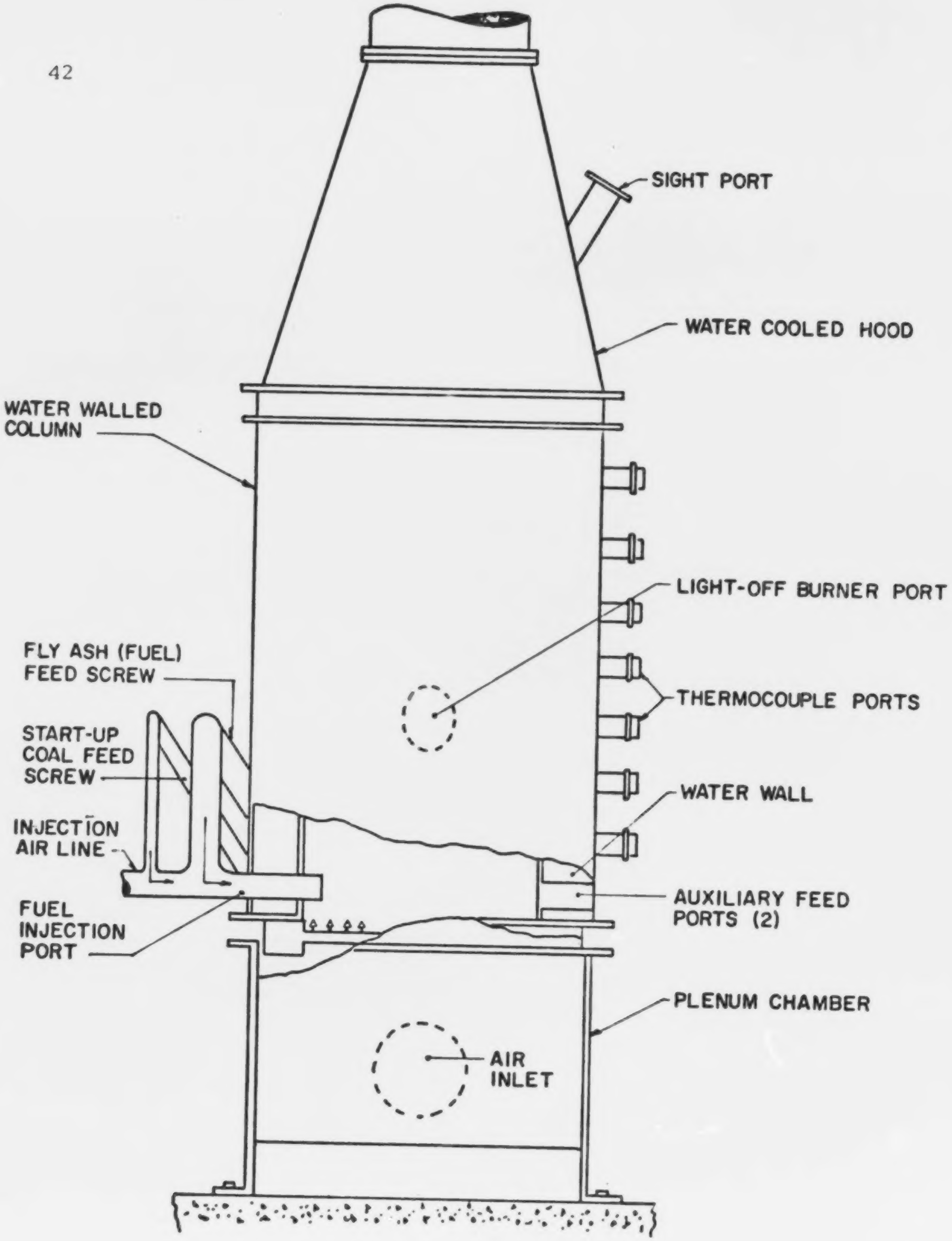


FIGURE 4 FLUIDIZED-BED COLUMN (FBC) CONSTRUCTION DETAIL-SIDE VIEW

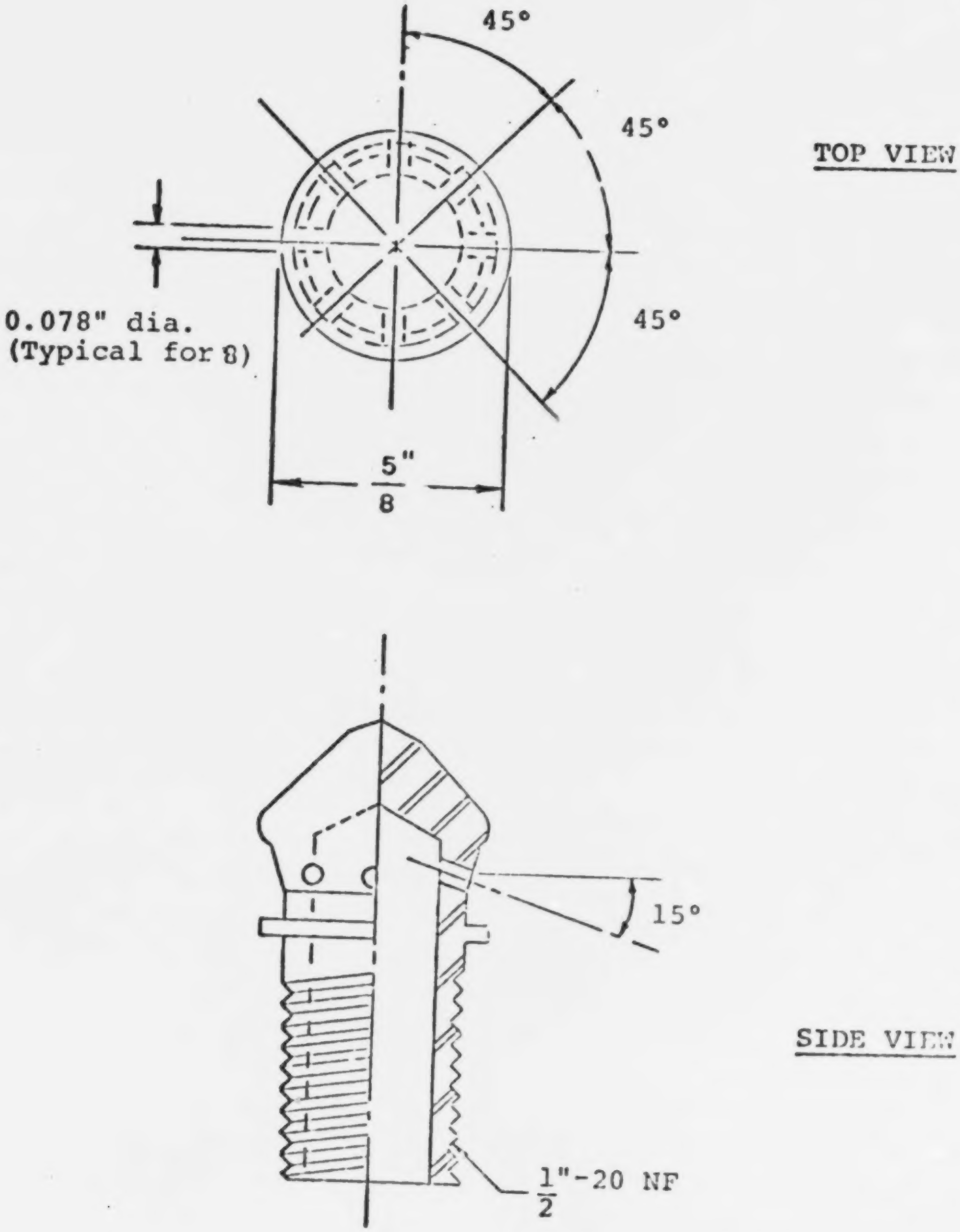


FIGURE 5. AIR DISTRIBUTION GRID BUTTON

a heat release of approximately $2 \times 10^6$ Btu/hr. These are both in excess of the actual operational values, since the FBC's performance as a fluidized-bed combustor deteriorates at firing rates much in excess of $1.2 \times 10^6$ Btu/hr. The water walls of the FBC will remove on the order of $0.5 \times 10^6$ Btu/hr from the fluidized-bed. As will be shown, it is not possible to sustain the fly-ash combustion process with that level of heat loss. It was therefore necessary to reduce the energy removed by the water-walls, and the FBC was insulated internally, as shown in the partial cross-section of Figure 6. The water-cooled hood was insulated in a similar manner. The insulation consists of a sleeve, or liner, of ASTM 446, one of the most refractory steels, backed with 1" of Kaowool, a refractory insulation. With the insulated sleeve, the FBC has an internal cross section of 0.86 $ft^2$. With the insulation in place, high bed temperatures, in the 1800°F to 2100°F range, could be achieved even with relatively deep beds, up to a 22" static depth.

The temperature of the bed could be controlled by the use of cooling surface in the form of bayonet tubes, also shown in Figure 6. The temperature of the bed could be decreased by lowering one or more bayonets. Four, fully inserted, would provide 4 $ft^2$ of heat transfer surface.

The bed material consisted of sintered coal ash, crushed and screened to either -16 +22 or -8 +16 U.S. Standard sieve size. The size was selected on the basis of the air rates to be used in a particular test; the largest particles for the highest air rates.

To begin a test, the bed is heated to coal ignition temperatures with a premix gas burner flame directed downward onto the bed as shown in Figure 3. The ignition procedure

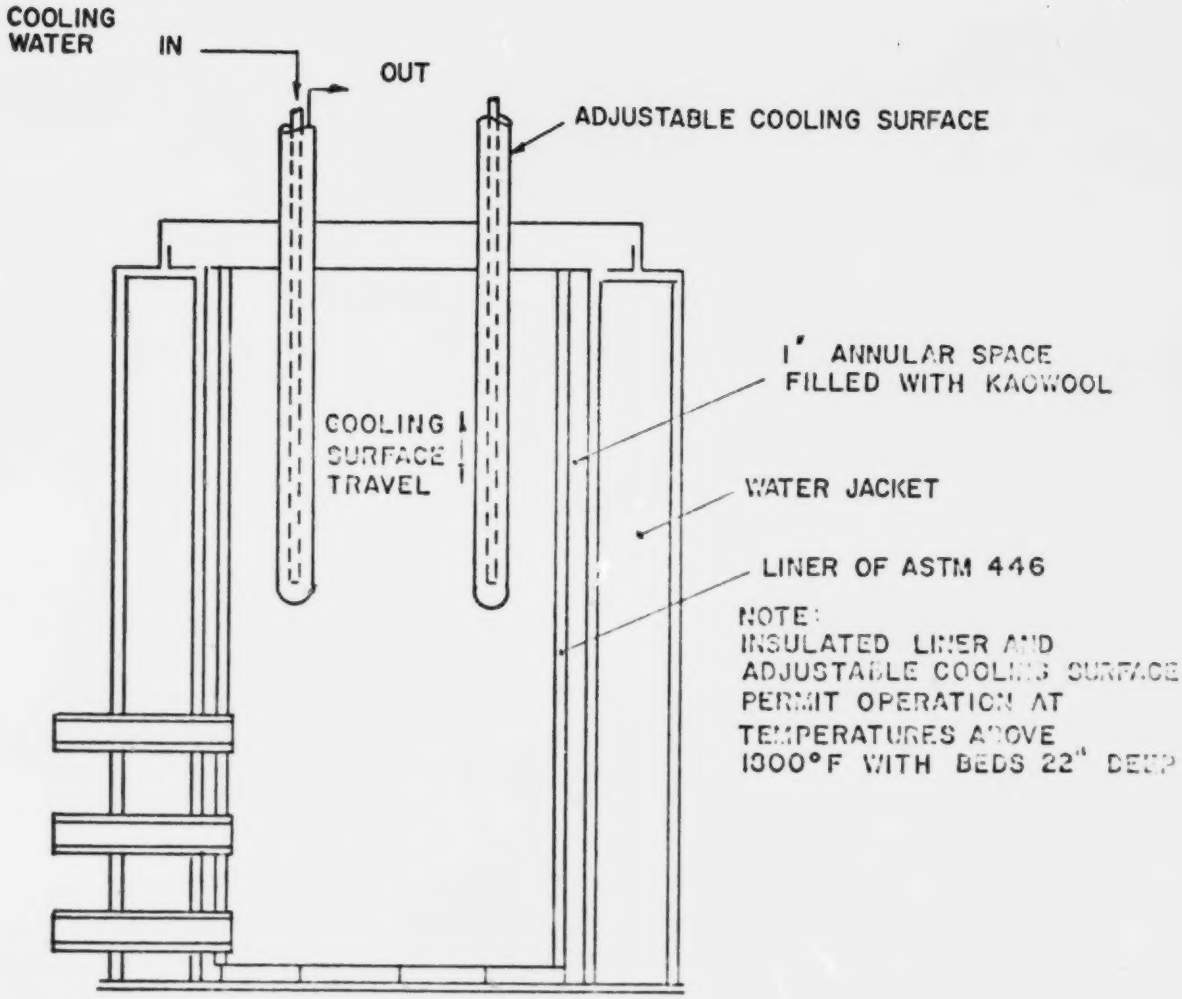


FIGURE 6 SECTION THROUGH FLUIDIZED-BED COLUMN SHOWING INSULATED STEEL LINER AND ADJUSTABLE COOLING SURFACE

involves fluidizing the bed material with minimum air flow, raising the bed temperature to 800°F, then injecting coal until the combustion is self sustaining while raising the air rate. About 10 minutes of heating is required for ignition. After the bed is operating stably with coal, the feed of carbon containing fly ash may be initiated. Fly ash feed rate is then increased as coal feed rate is reduced toward zero.

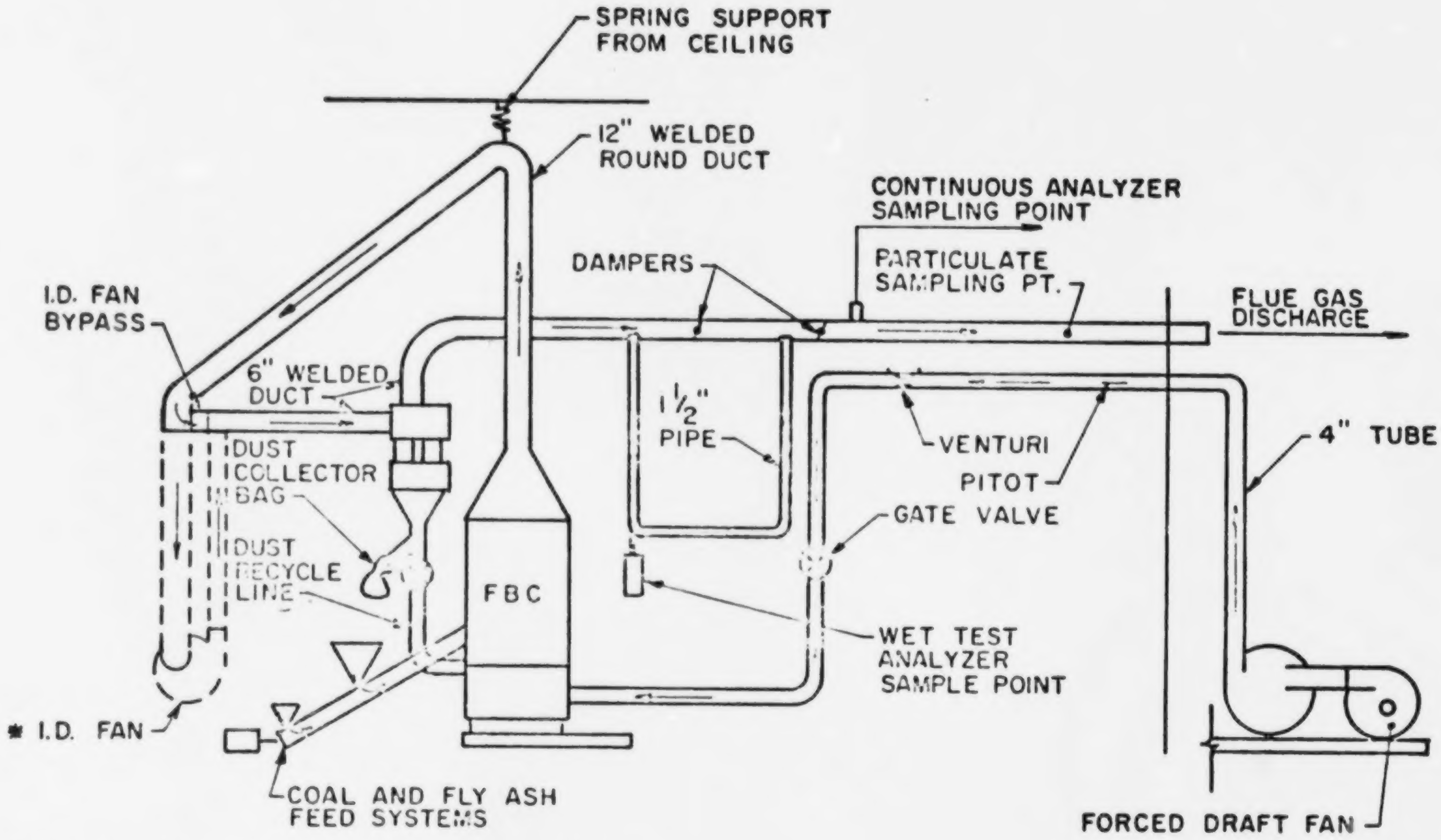
The entire FBC test system is shown schematically in Figure 7. Air is supplied to the plenum of the FBC by two blowers operating in series. Combustion products from the FBC pass through an induced draft fan (which may be bypassed), through a dust collector and on to atmosphere. The slanted duct between the FBC and the dust collector provides gas cooling without causing wall surface temperatures to fall below the dew point of the gas. A control damper may be used to provide a variable back pressure on the system. Fluidizing combustion air was monitored both by a pitot tube and a venturi meter located in the long entrance duct. A gate valve in the line was used to control airflow to the unit. The fuel feed rate was controlled by variable speed drives on the fly ash and coal feed screws.

Collected dust was discharged into bags and weighed. A dust recirculation system, indicated in Figure 7, was not used in this test series; neither was the induced draft fan.

## 5.2 Full-Scale Boiler, FBM

### 5.2.1 Boiler Module, FBM

The full-scale boiler module, designated the FBM, is a



* I.D. FAN NOT USED IN THIS TEST SERIES

FIGURE 7 SCHEMATIC OF FBC AIR AND EXHAUST GAS DUCTING SHOWING SAMPLING POINTS

coal-fired boiler capable of generating about 5,000 pounds of steam per hour at steam pressures up to 300 psig. A large amount of work has been conducted with this unit in past studies (1, 4).

The term FBM is used when referring to the test unit as a whole, which includes auxiliaries and appendages, and is also used when referring to that part of the whole being described here. Whether the whole or part is being discussed will be clear from the context. The term "primary cell" is also used when referring to the FBM (part) and "burnup-cell" is applied to the CBC. The FBM represents one half-cell of one of the original Multicell Fluidized-Bed Boiler concepts developed under contract with the Office of Coal Research. Two modules placed end-to-end would comprise one cell. A number of cells placed side by side without intervening insulation would make up a full-scale boiler.

A 250,000 pound per hour boiler uses about 25,000 pounds of coal per hour. Since the FBM is capable of firing 800 pounds of coal per hour, about 30 FBM-like modules would be required for the large boiler.

A cutaway sketch of the FBM is provided in Figure 8. A simulated Carbon-Burnup Cell, designated the CBC, added at the rear of the FBM, is also shown in Figure 8. The CBC is discussed in Section 5.2.2.

The FBM cross section (not including the CBC), is approximately 18 by 72 inches, roughly seven times the cross section of the FBC without the insulating liner. The bed is surrounded by vertical boiler tubes which extend from the two cross headers, below the grid plate, to the steam drum. No other tubes were placed in the bed for the tests discussed in this report. The boiler tubes are joined together by

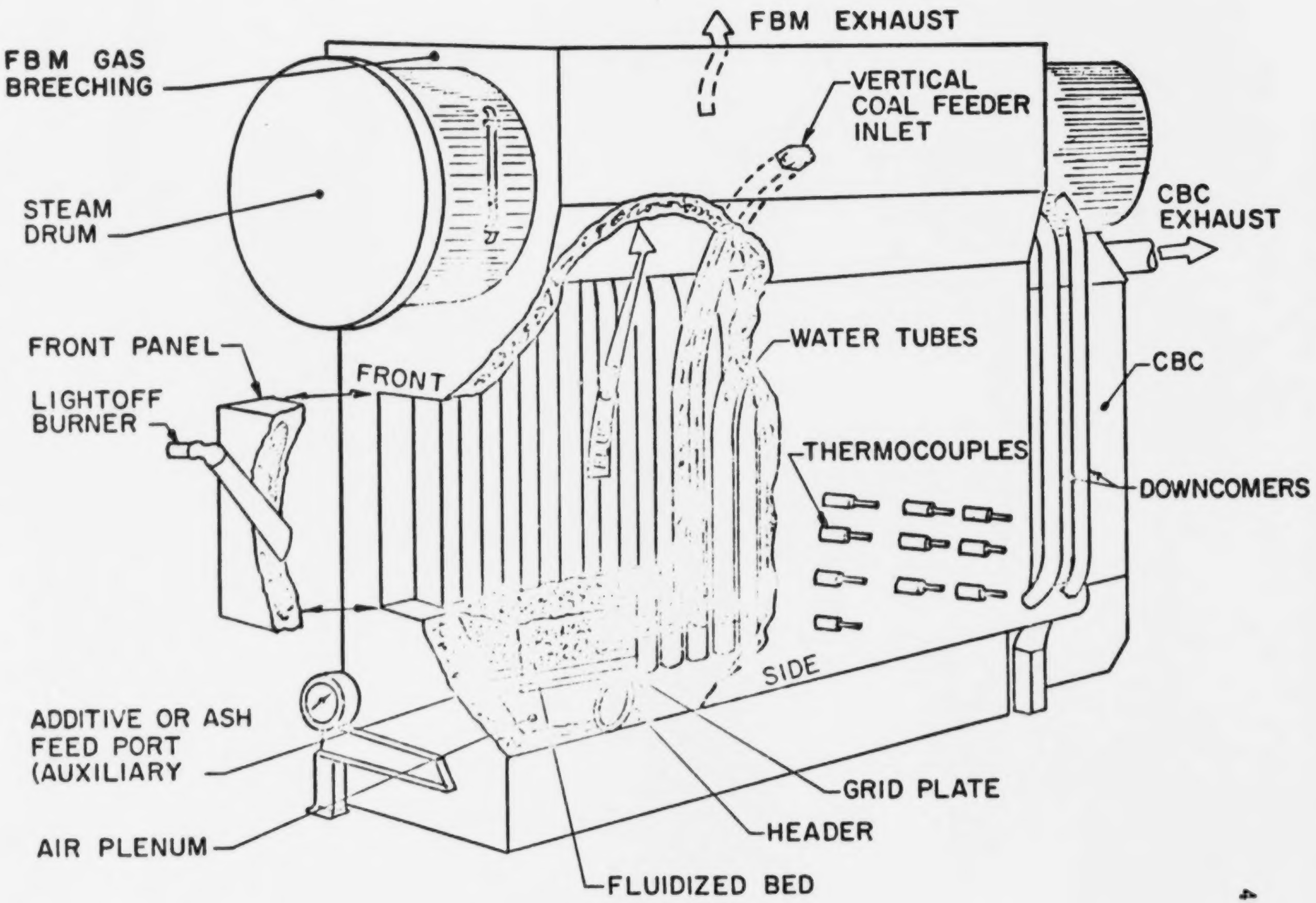


FIGURE 8 FLUIDIZED-BED MODULE (FBM) INTERNAL CONSTRUCTION

welded fins and are backed by insulation. The fins do not extend the full height of the tubes so that flue gas passes between the tubes at the top of the unit, around the steam drum and into the breeching.

The combustion space of the FBM is accessible through a water-cooled panel at the front of the unit. The panel contains a pre-mix gas burner used to fire the bed. The burner directs a flame downward onto the front of the bed. From the plenum at the base of the unit, combustion air is directed upward through the grid and then through the bed. The grid consists of a mild steel plate containing buttons of the same spacing and design used in the FBC. The bed material used in the FBM tests was sintered ash, -8 +20 U.S. Standard mesh. The static bed depth may be varied from 6 to over 30 inches although the useful range is narrower. Thermocouples were mounted throughout the bed as shown in Figure 8. Detailed specifications of the FBM are presented in Appendix B.

For this test series a vertical coal feeder, shown in Figures 8 and 9, and discussed in Section 7.2 was utilized although a number of other designs have been considered (4). Sorbent and, in some tests, fly ash were fed via four horizontal ports that penetrated a side wall of the unit, through holes cut in the fins, just above the air distributor. (See Figure 9). The ports -- numbered 1, 2, 3 & 4 from the front of the FBM -- were spaced along the wall opposite the thermocouples shown on Figure 8. The ports extend 1-1/2" into the bed and end with a deflector to impart an initial downward path to the fine powders fed.

In operation, the bed is raised to the ignition point of coal by use of the gas burner. Combustion of the coal

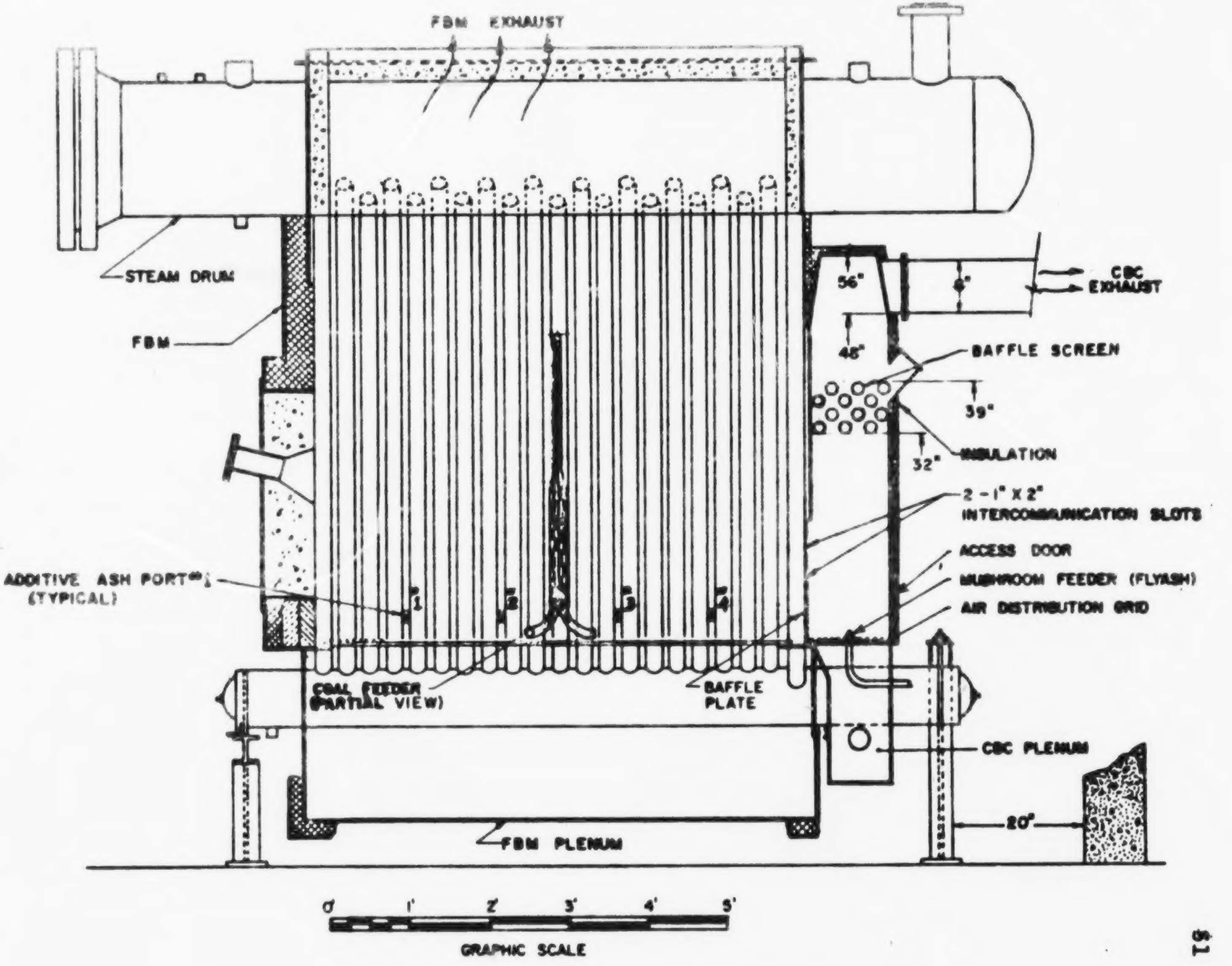


FIGURE 9. SECTION THROUGH FLUIDIZED BED MODULE (FBM) AND SIMULATED CARBON BURNUP CELL (CBC)

begins in the vicinity of the light-off burner and propagates rapidly throughout the bed.

The energy not absorbed by the waterwalls leaves the FBM as hot products of combustion. Two water-cooled tube banks and an air heater were installed in the breeching beyond the FBM to absorb some of this energy.

A schematic drawing of the FBM test system is shown in Figure 10. Air from an external forced-draft fan passes through the air preheater and into the FBM plenum. Coal feed is controlled by the speed of a screw feeder which drops the coal into a pneumatic feed tube at the injection port. Sorbent materials were also screw fed to another pneumatic injection line at a rate controlled by a variable speed screw drive. Fly ash transport is pneumatic. A wear resistant star feeder provides the pressure seal between the dust collector hopper, maintained under negative pressure, and the positive pressure transport line. Although the details can not be shown in Figure 10 primary cell fly ash can be discharged to a storage hopper, sent to the CBC or be returned to the FBM. Generally in this test series the collected fly ash was fed to the CBC. Flue gas from the FBM passes across the first gas cooler above the unit to reduce temperature before the gas enters the air preheater. As the flue gas passes through the air preheater, a portion of the fly ash drops out and is collected in the hopper shown. The bulk of the fly ash is removed by a multi-cone collector downstream of the air heater. During recirculation, the ash knocked down by the preheater is screw fed into the dust collector hopper. From the collector the gas flows through a long duct to an induced draft fan and then to atmosphere. A damper is provided in the ducting to control pressure in

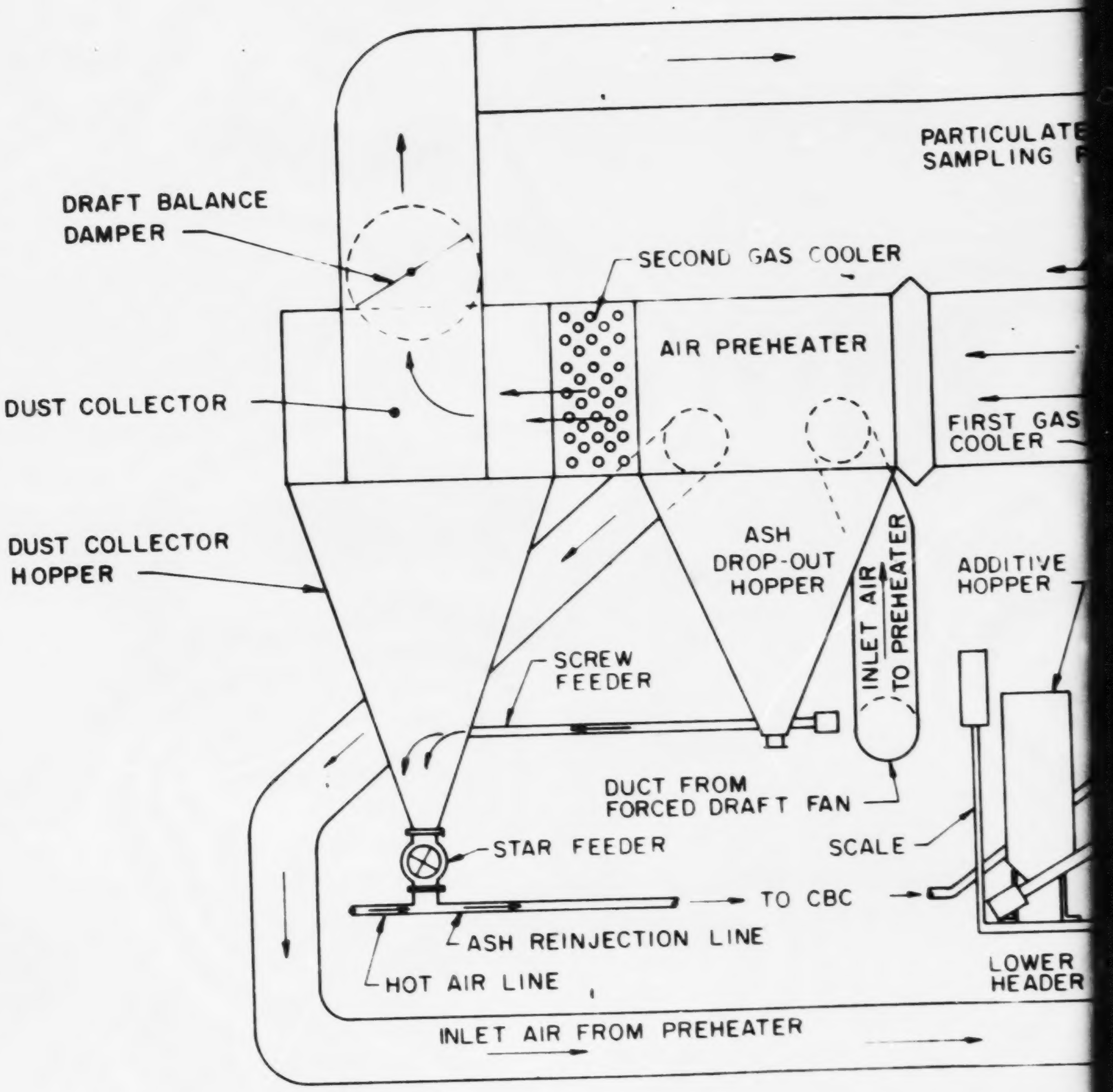
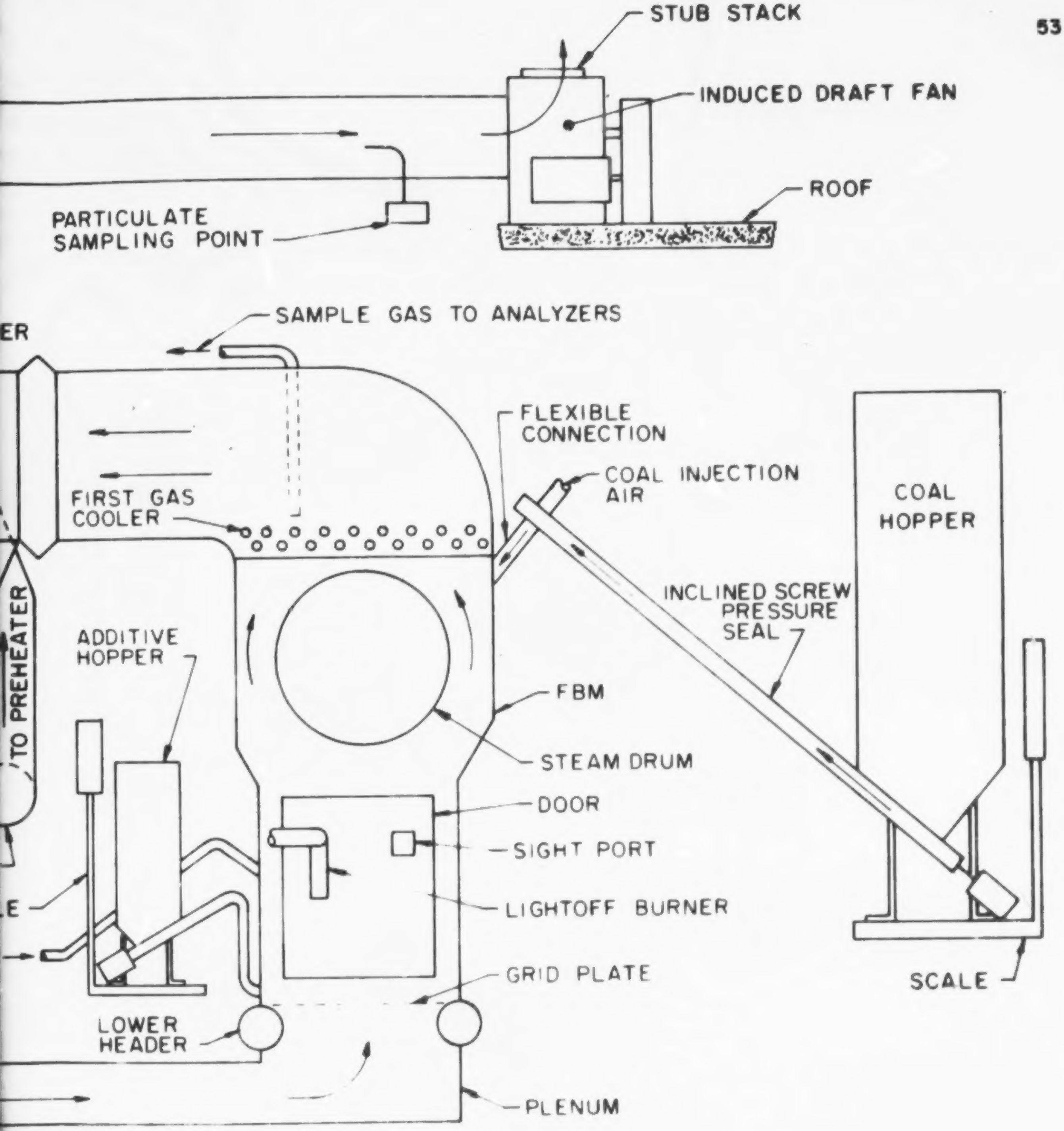


FIGURE 10 SCHEMATIC OF FBM TEST SYSTEM SHOWING



TEM SHOWING VARIOUS SUBSYSTEMS

the combustion chamber. The system is operable without the induced draft fan but is not usually run pressurized.

### 5.2.2 The Simulated Carbon-Burnup Cell, CBC

A demonstration of the Carbon-Burnup Cell concept, operating on a continuous basis, was required. Although it would have been possible to operate the FBM and FBC in tandem (FBM firing coal, the FBC firing the FBM's fly ash) this might not have been considered an adequate demonstration of the efficacy of the Carbon-Burnup Cell concept. Building a test system with an actual Carbon-Burnup Cell, while providing the most effective demonstration, was not possible. It was therefore decided to modify the FBM by adding an appropriately sized simulated Carbon-Burnup Cell. This device was designated the CBC.*

The CBC was to operate continously, burning the carbon containing fly ash produced in the FBM while the FBM was fired with crushed coal. The CBC was to operate at a higher bed temperature that the FBM, but -- since bed material could circulate between the two sections -- they shared a common bed. No precendents existed for the design of such a system, accordingly the apparatus was modified as the program progressed.

The CBC consisted of a chamber with a rectangular cross-section, 10-5/8" x 15-5/8". From the grid level, considered the base line for all vertical dimensions, to the roof of the CBC measured 56" but to the gas outlet the height of the CBC was only 48". These dimensions are shown in Figure 9.

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*The FBC and CBC were both simulated Carbon-Burnup Cells because they did not have cold walls. To use cold walls the cell must be large. In a small unit the chamber walls must be insulated. When refractory walls are used the ash attacks the refractory and sintered slag accumulations can grow. When insulated, oxidation-resistant metal walls are used, as in these tests, slag does not grow but the walls are destroyed in only a few hundred hours of operation.

Figure 9 also shows why these inadequate height dimensions were used. The only way to make the CBC an integral part of the FBM without cutting pressure parts was to locate it at the back and under the steam drum. The maximum width was set by the down-comer spacing and the depth, by the location of one of the building's structural walls. Except for height, the dimensions were appropriate for the expected firing rate (500,000 to 750,000 Btu/hr).

The decision to avoid cutting the drum or removing some side wall tubes* was an error as the difficulties encountered with bed material loss (discussed below) indicated.

The uncooled, insulated walls of the CBC were fabricated of 27% chromium steel, designated ASTM 446. This metal has the lowest thermal coefficient of expansion of any stainless steel and is able to resist oxidation at 2,000°F. The grid plate was of the same design as that in the FBM.

As shown in Figure 9 the exhaust was an 8" I.D. horizontal duct. This duct extended horizontally under the steam drum, through the laboratory wall and then turned vertically to a dust collector installed on the laboratory roof. Dust collected here could be sent to a storage hopper, routed back to the FBM via the four point injector or returned to the CBC.

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*In the work which followed that reported here the FBM was essentially remade, in place, by a very small, very skillful ASME certified welder.

Fuel for the CBC was the fly ash generated by the FBM. Coal was fed along with the fly ash when the fly ash alone could not maintain the desired 1900°F to 2100°F bed temperature level. Fly ash from the CBC's own collector was also fed back into the CBC during most tests described in this report. All fuel entered the unit through one port.

The fuel injector initially employed in the CBC was a square, horizontal tube jutting into the unit almost identical to that used in the FBC. However, this design was soon changed to a "mushroom" feeder such as that shown in Figure 11 and positioned as in Figure 9.

Fluidizing/combustion air is supplied via the air heater from the main FBM air supply. Although the design of the mushroom feeder makes it capable of supplying the bulk of the CBC's fluidizing air the plenum and air distribution grid were always used.

The FBM and CBC share a common wall which consists of the FBM's 2 inch O.D. back-wall tubes with intervening fins. Portions of these fins were removed to enable bed material to circulate between the two sections. As shown in Figure 12 five slots were cut out of the fins 18" high and 1" wide. A 2" O.D. boiler tube was between each pair of slots. The part of each of these tubes facing the CBC was insulated by a semicircle of insulation held in place by a thin metal shield. The slots provided 90 square inches of open area in the wall between the two bed regions.

The open area between the two sections, provided by

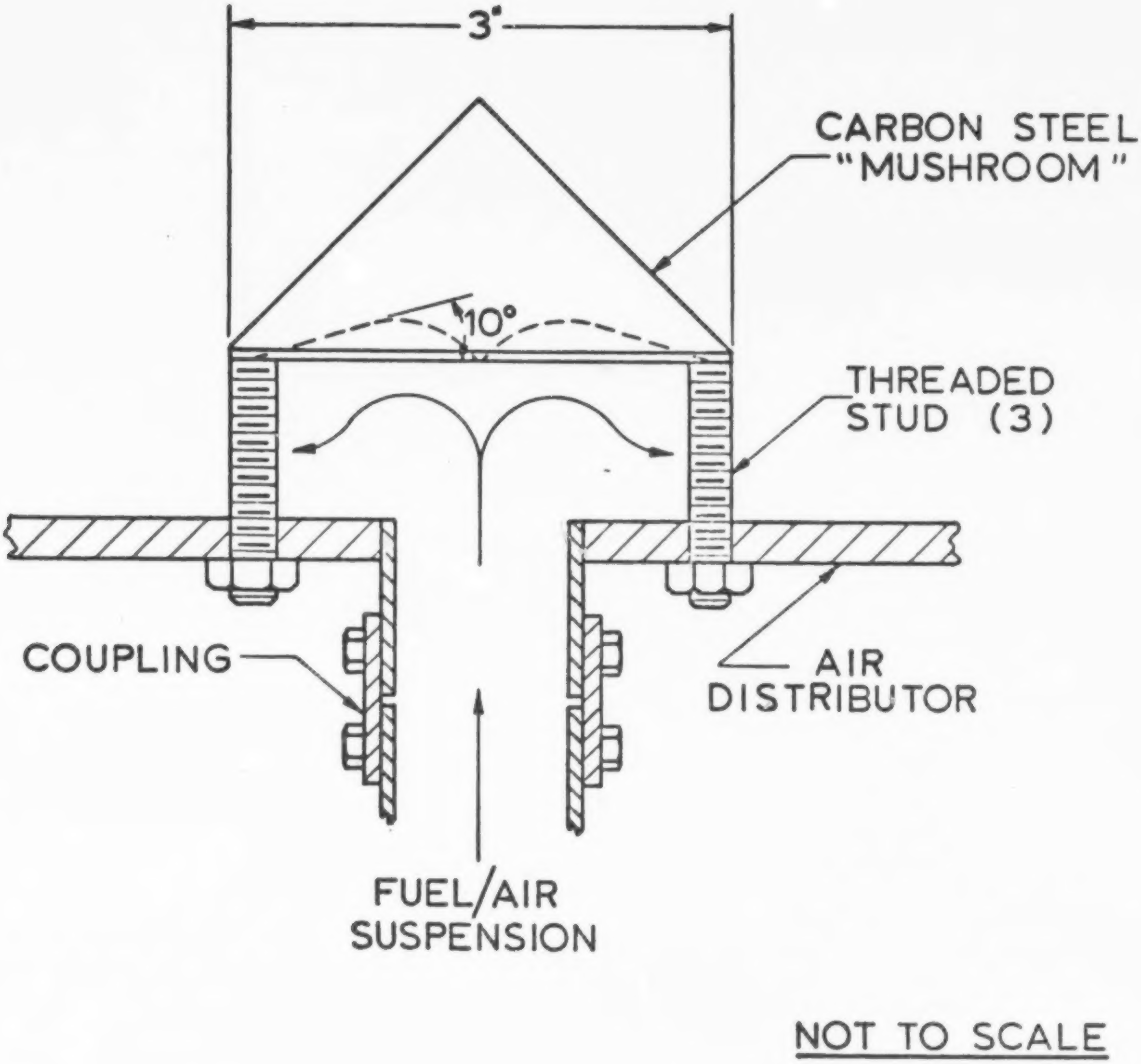


FIGURE 11 MUSHROOM FEEDER FOR FLUIDIZED-BED COMBUSTOR

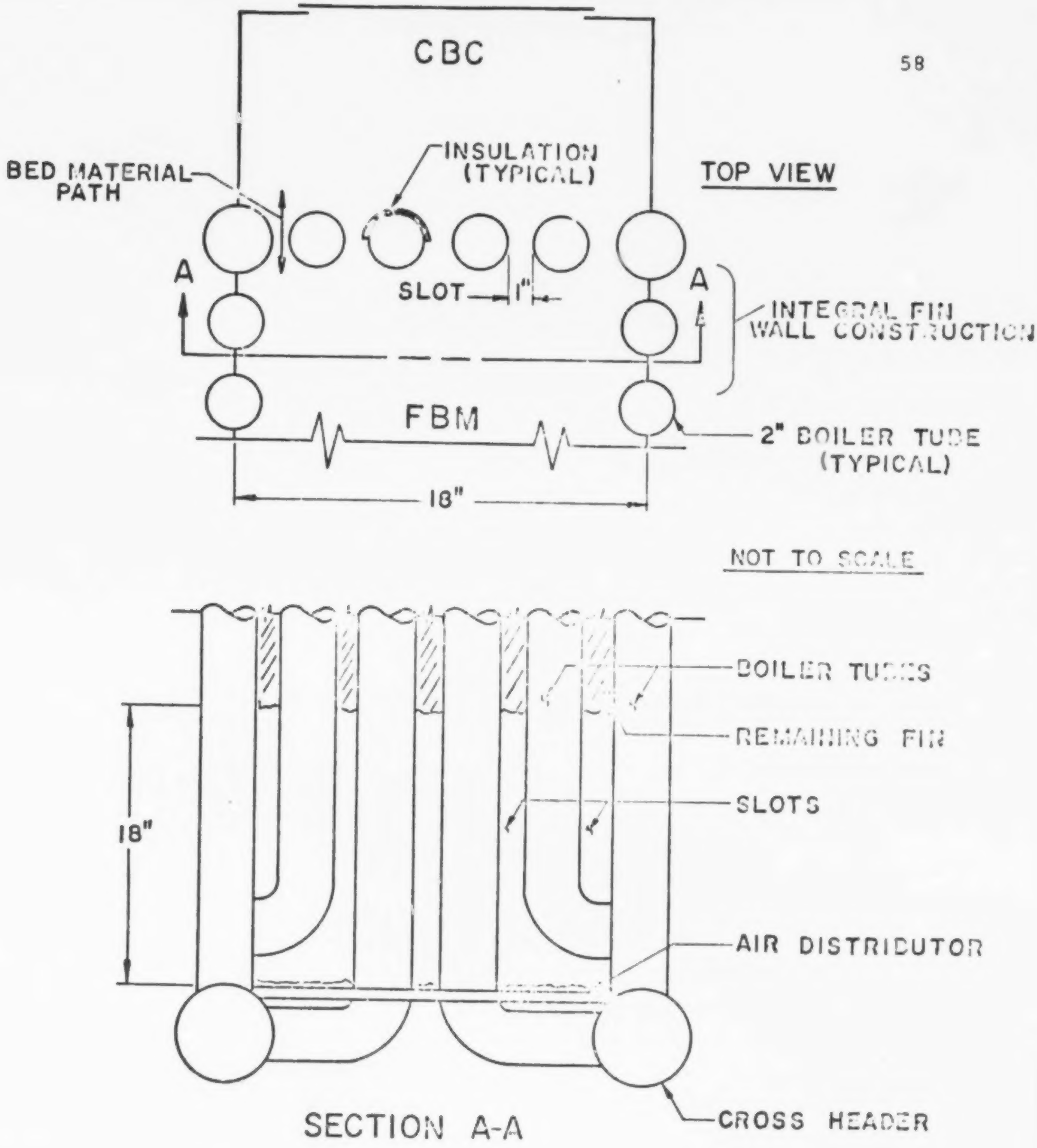


FIGURE 12 DETAILS OF WALL BETWEEN FBM AND CBC

removing the fins, was found to be excessive for the volume of bed material in the CBC. Bed material circulates freely between the two sections and this interchange provides a significant heat sink for the CBC. Each pound of bed particles which enter the CBC at 1550°F and leave at 2050°F removes 120 Btu from the CBC.

The 90 suare inches of area provided so great a rate of bed material exchange that the desired temperature difference, 500°F between the bed material in the FBM and the bed material in the CBC, could not be achieved. Thus, in order to reduce the rate of bed material circulation, the open area in the common wall was reduced in three stages: to 45 square inches, to 4 square inches, and then finally to 2 square inches. The effects of these changes are discussed in Section 7.4. The reduction of slot area was achieved by placing a carbon-steel baffle on the CBC side of the common wall of boiler tubes thus blocking the open area left by the removal of the fins. Holes cut in the baffle provided the desired opening.

The initial test results with the CBC as part of the FBM were poor because of the initial injector design, the excessive open area between the two sections and finally because of the insufficient freeboard in the CBC chamber. The horizontal exhaust from the CBC was simply too close to the top of the expanded bed to allow bed particles, "splashed" up into the freeboard, to fall back into the bed before being swept out by the gas. Cooler bed material from the FBM, would flow into the CBC to replace the material being lost at a higher temperature. To reduce this energy loss two designs for "particle knockouts" were tested. The basic design concept was based on information obtained in

April 1970 from BCURA Industrial Laboratories. Figure 8 shows how these two knockouts were placed in the CBC.

The first design consisted of horizontal triangular array (4 rows) of 1" tubes connected by U-bends to form a single continuous water circuit. Water flowed into the tube at the bottom and out at the top. Thermometers installed in the inlet and outlet permitted the temperature rise to be determined. A total of 5.2 square feet of heat exchange surface was provided by this knockout screen. The effect of this heat exchange surface is discussed in Section 7.6.

The second design consisted of a similar array of uncooled rods fabricated of ASTM 446. This screen, as well as the first, was installed within the CBC extending from the 32" level to the 39" level. The knockouts are discussed further in Section 7.5.

The test procedures for both the FBC and FBM operations involved igniting the bed and stabilizing the combustion at the desired bed temperature until steady-state conditions prevailed. Steady-state was assumed when the bed temperature was constant and the gas composition analyzers indicated constant flue gas composition.

## 5.3 Instrumentation

Emissions of sulfur dioxide, nitric oxide and hydrocarbons were monitored continuously. Infrared analyzers (Beckman 215) were used to monitor sulfur fioxide and nitric oxide. Hydrocarbons were detected with a flame ionization analyzer (Beckman 109A) using methane as the reference gas. The signal output of each of these units was displayed on strip chart recorders.

The gas transfer system used with these analyzers is sketched in Figure 13. The system permitted rechecking of calibrations on any of the three units at any time during a test by switching from sample gas to reference and zero gases at the rotameter valves. The sample gas was drawn from the flue gas stream through a sintered stainless steel filter and conditioned to remove water. The sample gas was again filtered before entry into the analyzers to prevent possible contamination of the optical cells and the hydrogen burner.

The FBC gas sample was drawn into the instrument room through a heated 1/2" O.D. tube from the FBC flue gas discharge duct, shown in Figure 8 which passed directly overhead.

In sampling the FBM flue gas, special precautions were necessary because of the possibility of infiltration of dilution air in the duct above the unit. Also, the poor instrument response which would result from drawing a small sample a long distance (approximately 60 feet) from unit to instrument room was undesirable. Thus a system was devised to draw a large gas sample from the FBM, just above the first gas cooler (See Figure 10), pass it through a dust collector, and then through a loop above the instrument room. The sample tube was a 3" pipe with sections screw-fitted and then seal welded. The system was driven with an I.D. fan located at the discharge to atmosphere. Where the sample line passed over the instrument room, a small sample was withdrawn via a 1/2" O.D. tube. A schematic drawing of the system is shown in Figure 14.

The CBC gas sampling system was similar in design to that for the FBM. The CBC sample line was a 2" pipe located in the exhaust duct upstream of the dust collector, as shown

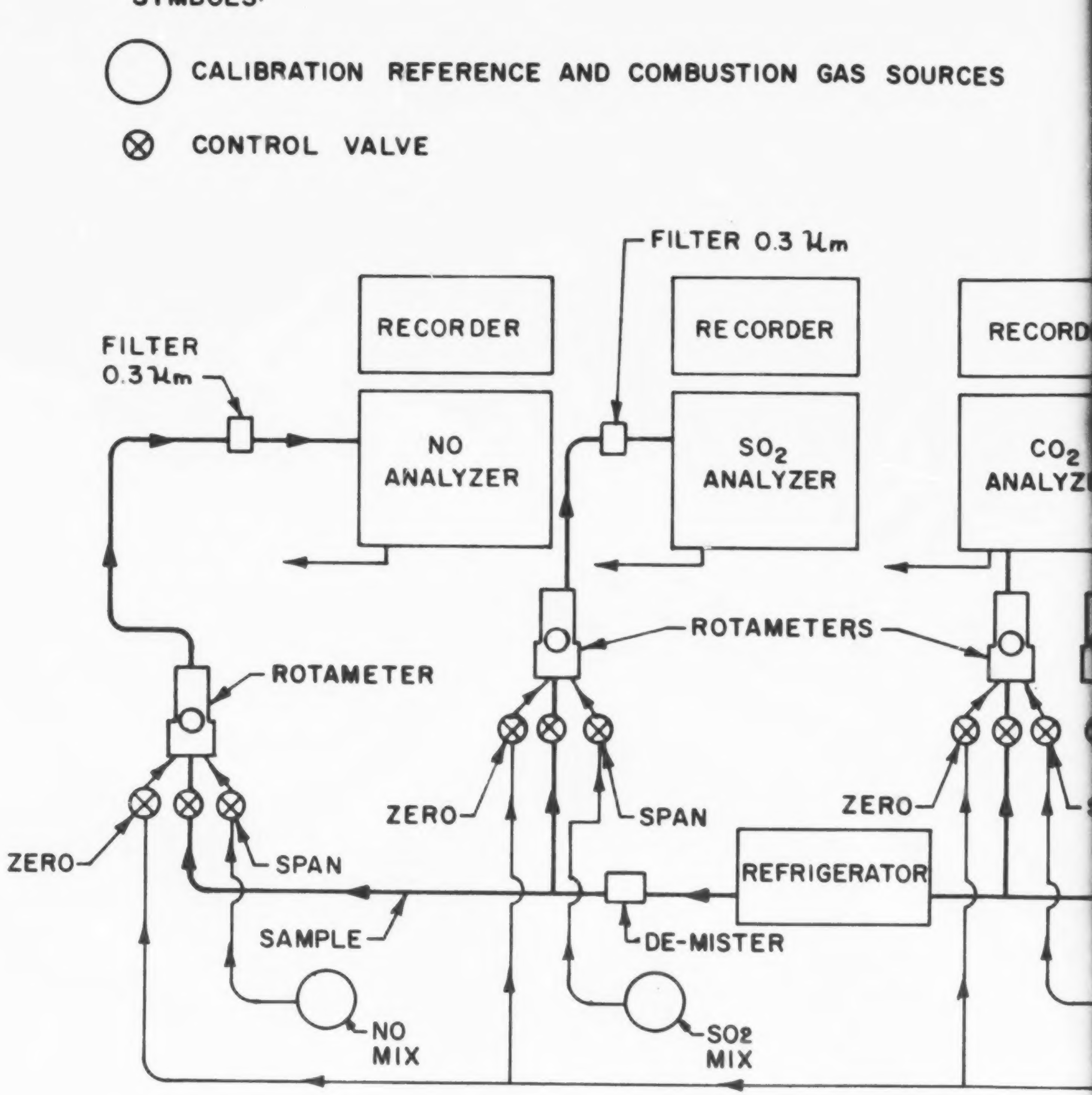
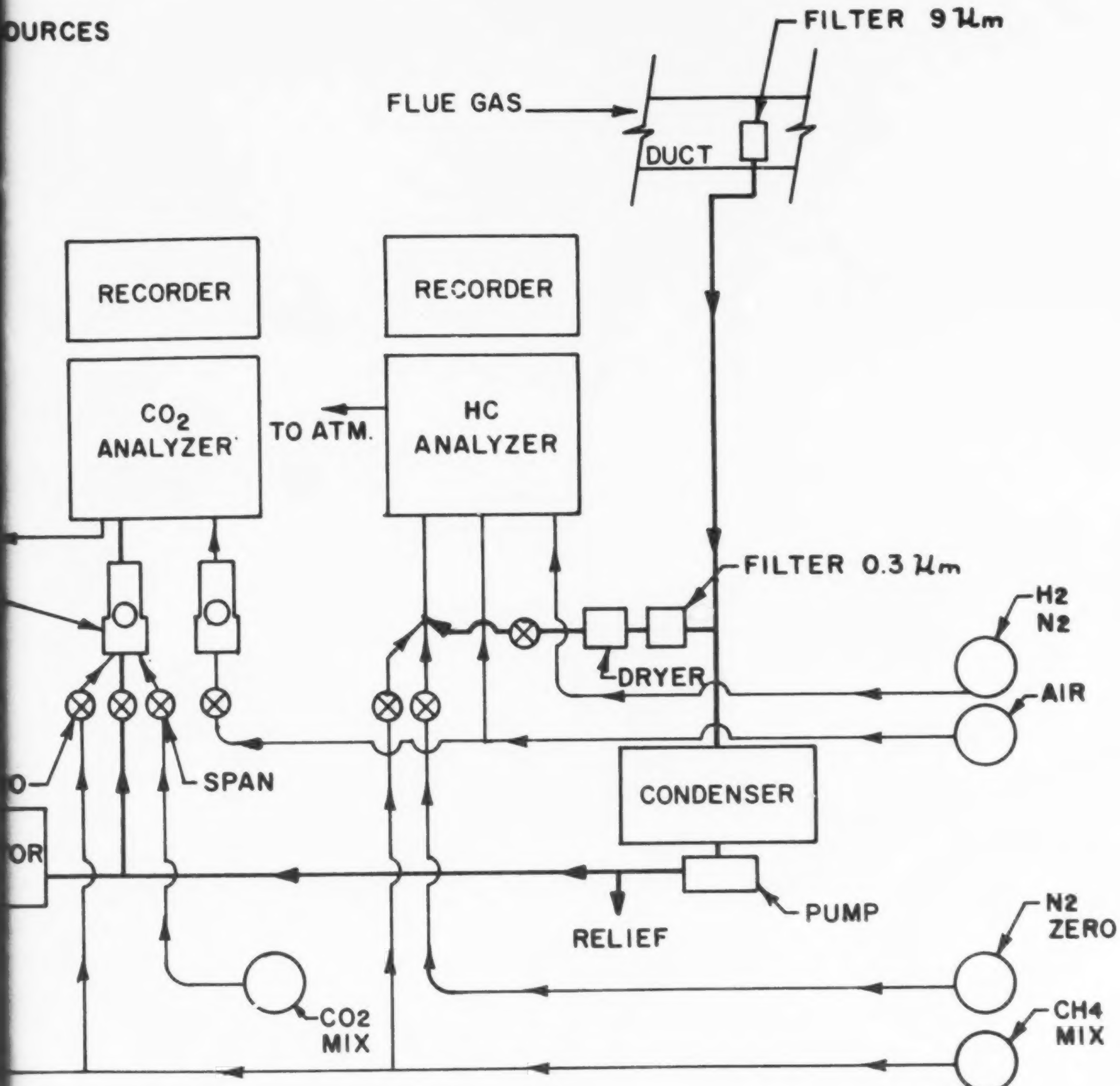


FIGURE 13 SCHEMATIC OF GAS TRANSFER SYSTEM FOR CONTINUOU
NITRIC OXIDE, CARBON DIOXIDE AND HYDROCARBONS

OURCES



OR CONTINUOUS MONITORING OF SULFUR DIOXIDE,
HYDROCARBONS

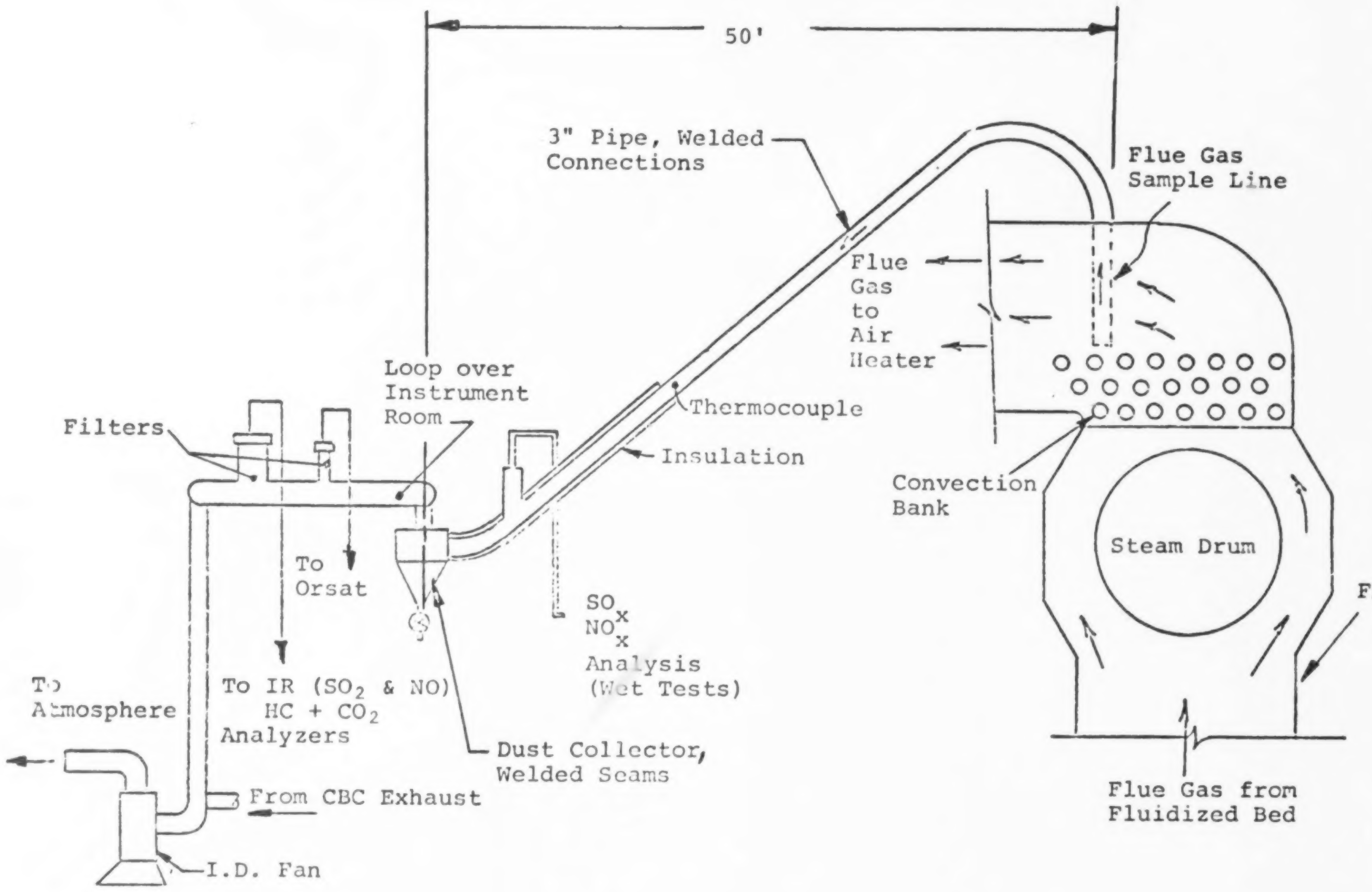

50'
3" Pipe, Welded Connections
Flue Gas Sample Line
Flue Gas to Air Heater
Thermocouple
Insulation
Convection Bank
Steam Drum
FBM
Flue Gas from Fluidized Bed
Loop over Instrument Room
Filters
To Orsat
$SO_x$ $NO_x$ Analysis (Wet Tests)
To Atmosphere
To IR ($SO_2$ & NO) HC + $CO_2$ Analyzers
Dust Collector, Welded Seams
From CBC Exhaust
I.D. Fan

FIGURE 14. SCHEMATIC OF THE FBM GAS SAMPLING SYSTEM

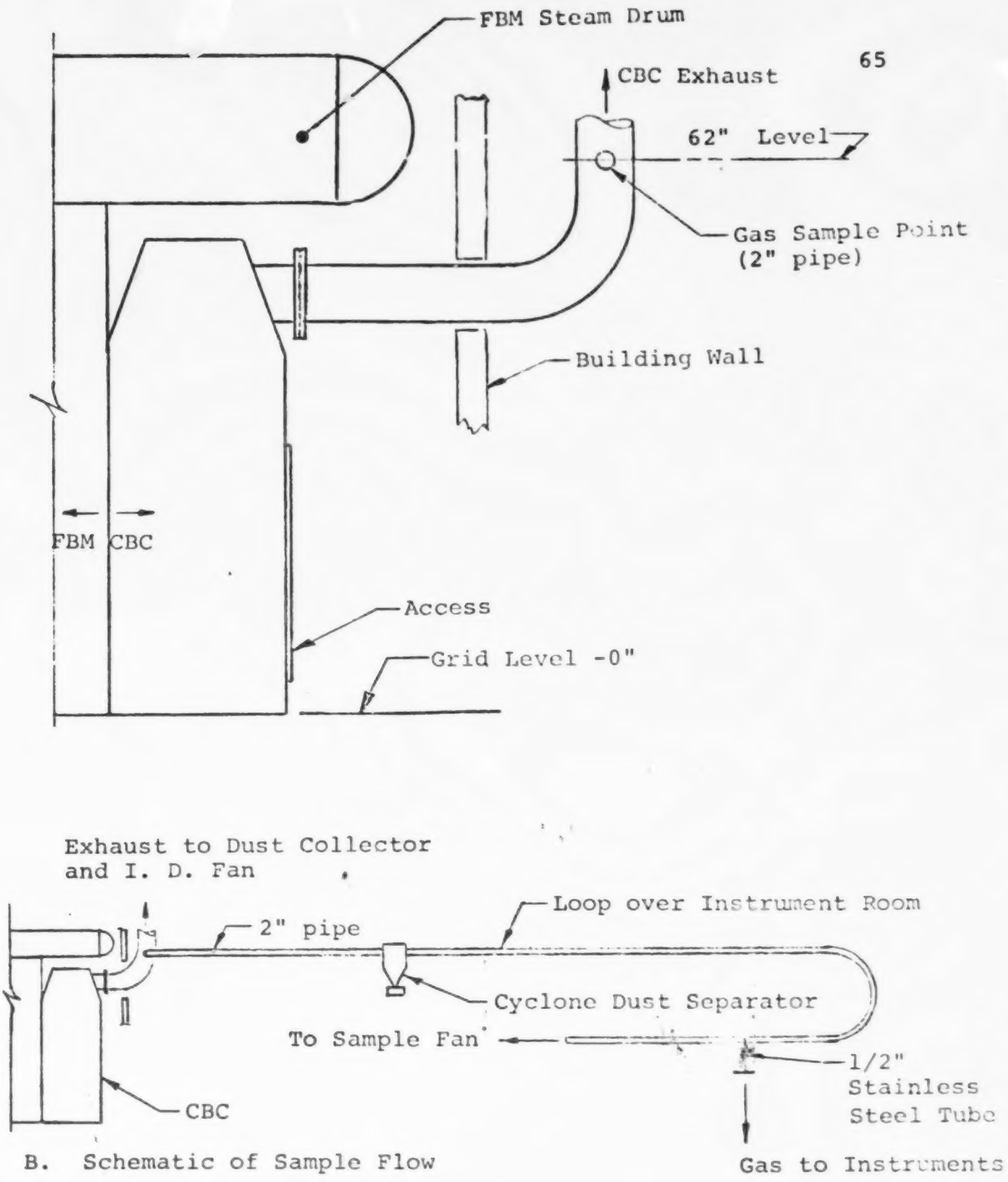
in Figure 15.

When the FBM was operated, it was not possible to simultaneously analyze the gas from the primary cell and the burnup cell. A pair of ball valves permitted a sample to be drawn from the pipes above the instrument room carrying the FBM and CBC sample streams. The time that the analyzers spent on each stream varied but was generally about 15 minutes.

Particulate emissions were monitored with the isokinetic probe system described in Reference 1. The probe design permits equalization of the sampling velocity with the stream velocity. Locations of particulate sampling points in the FBC, FBM, and CBC test systems are indicated in Figures 7, 14, and 15, respectively.

Carbon dioxide was continuously monitored, using a self-referencing thermal conductivity analyzer (Beckman Model 7C). The reference and span calibration gases contain 16% $CO_2$. The zero reference gas is purified air. Instrument cabinet temperature is controlled at $130^{\circ}F$. The $CO_2$ analyzer was checked using a conventional Orsat analyzer which also determined $O_2$ and CO.

Bailey oxygen analyzers (Type OC1530A) were used as operator guides to indicate oxygen concentration in the FBM and CBC flue gases. During a test period, the air input rate was held constant and the fuel rate adjusted to maintain the oxygen concentration at the desired value. The Bailey instruments have been calibrated periodically with known mixtures of $O_2$, $N_2$, and $CO_2$ and found to be very reliable. The flue gas oxygen was also verified, using the standard Orsat technique. When the FBM was operated, a separate oxygen



B. Schematic of Sample Flow

FIGURE 15. SCHEMATIC OF CBC GAS SAMPLING SYSTEM

analyzer was connected to the primary cell and the CBC.

Temperatures in the bed and at various other points in the system were recorded on a Honeywell Multi-point Recorder. A multiple switch panel was used to connect the recorder to either the FBC or FBM systems, as required. Locations of thermocouples in the systems are indicated in Appendices A and B.

The infrared analyzers and the hydrocarbon analyzer were calibrated with gas mixtures supplied by vendors. The concentration of the active components in the calibration gases were checked after delivery to the laboratory. The methane mixture was analyzed by the National Bureau of Standards - a report is shown in Reference 1. This gas, containing 1265 ppm $CH_4$, was used to calibrate a second methane mixture before the first was depleted.

The sulfur dioxide calibration gas was analyzed with a peroxide absorption train. A gas concentration of 2650 ppm was used in the program. Analysis of the nitric oxide calibration gas is also given in Reference 1.

The output signal of the infrared sulfur dioxide analyzer varies in a nonlinear manner with $SO_2$ concentration. The calibration curve provided with the instrument was checked by precision dilution of the known calibration gas. The curve was found to be correct except for a slight deviation at the low end of the range. The calibration curve and check points are given in Reference 1.

The calibration curve was used without correction since the deviation is not more than 1% of full scale.

The calibration curve for the nitric oxide infrared analyzer is given in Reference 1. The contribution of water vapor to the signal output is significant with this analyzer. The water vapor correction determined by the supplier (180 ppm) was checked by testing a dry gas in the analyzer for comparison with a moist sample. A correction of 220 ppm was noted and incorporated in the data reduction. The range of this unit is 0-1000 ppm NO.

## 5.4 Materials

Three materials were used in the test program: fly ash, coal and limestone. The sources of each are discussed below.

### 5.4.1 Fly ash for Tests in the FBC

Fly ash having the properties desired for this test series was not easy to obtain. The two properties required were a coarse size consist and a relatively high carbon content.

An analysis of the fly ash from several of the nearby utility power stations, showed their fly ash to be unsuitable. Interestingly a few samples had carbon contents in excess of 30%, but the particle size distribution was considered too fine; the utility plants all used pulverized coal.

Stoker fired steam plants serve most of the federal facilities in the Washington area; their fly ash was of the proper type. Unfortunately the ash required for these tests had to be dry and free of bottom ash and most plants either wet the fly ash or mixed it with bottom ash as it entered the transport system to the ash silo.

The search for fly ash took us to the prisons of the State of Maryland and to the Glidden's Titanium Dioxide plant in Baltimore. Both had stoker fired boilers. The prison officials thought our request for fly ash was rather odd but were willing to cooperate if we could obtain the proper clearances to bring a 6000 gallon solids tanker onto the prison grounds. Glidden was chosen.

A tanker was leased and parked adjacent to the fly ash hopper of the mechanical collector on one of Glidden's boilers. Pope, Evans and Robbins personnel constructed a system for withdrawing the fly ash from the hopper, which was negative pressure, and over a period of seven days the tanker was filled.

Unfortunately the material which arrived in the tanker was low in carbon ( ≈ 30%), not at all like the initial samples ( ≈ 45% carbon). Investigation revealed that the plant personnel in an attempt to be helpful had increased the load of the boiler filling the tanker, so as to fill it faster, and had thereby improved the units combustion efficiency.

The steam plant serving the U. S. Capitol, the Supreme Court and the Library of Congress agreed to operate their ash disposal system in a manner which would allow Pope, Evans and Robbins to obtain fly ash free of bottom ash. They would store fly ash while the ash silo was emptied, then transfer fly ash and store bottom ash. By this means a few hundred pounds of fly ash could be obtained during each trip, enough for a few tests. Fly ash obtained by this method constituted about 50% of the material used in the FBC statistical test program.

After the CBC had been added to the FBM so that the FBM could be operated, coal was burned in the FBM and the fly ash stored. This material was also used in the FBC.

The test data (Appendix E) do not indicate the source of fly ash for each test since this was not considered relevent by the statistical consultant. Capitol Power Plant fly ash was used in tests C-304 through C-312. FBM fly ash was used in tests C-313 through C-320. For tests C-301 through C-303, (these data were not used in the model) fly ash from PER laboratory's waste pile was used. No attempt was made to interpret FBC data for each fly ash source separately. The performance model provuded with all of the data indicates that this decision was valid.

### 5.4.2 Coal for FBC and FBM Tests

Most of the coal used in the tests reported here was from the same mine as that used in Reference 1. This coal is an East Onio Pittsburgh No. 8 Seam coal designated "Perfect 8". The high ash content indicates the coal was unwashed. The coal had the following analysis:

| Constituent | Ultimate Analysis, Wt. % (As Received) |
|---|---|
| Carbon | 68.35 |
| Hydrogen | 4.57 |
| Oxygen | 5.48 |
| Sulfur | 3.30 |
| Nitrogen | 2.50 |
| Moisture | 2.30 |
| Ash | 13.50 |
| Total | 100.00 |
| Net heating value | 11,498 Btu/lb |

### 5.4.3 Limestone for FBC and FBM Tests

The limestone used in those tests in which fine limestone injection was investigated was the same No. 1359 stone used in the studies described in Reference 1.

# 6. PROCEDURES AND RESULTS: TESTS WITH THE FBC

## 6.1 General

Tests in the insulated FBC were intended to determine the effect of important operating variables on the combustion of fly ash in a Carbon-Burnup Cell. The FBC results were independent of any complicating factors resulting from the interchange of bed material between the burnup cell and primary cell. Tests in the FBM, in which bed material interchange occurred, and in which the feed rate of fly ash to the CBC could not be controlled, were useful for demonstration of, and insights into, the dynamics of the process. However, the FBM could not be used for the derivation of a performance model with the time and other resources available.

## 6.2 Statistical Design of Experiments

### 6.2.1 Perspective

Predictions, based on prior work in the field of fluidized-bed combustion, were made at the outset of the program. It was predicted that combustion efficiency, defined below, would be increased by increases in the following parameters:

1. Bed temperature.
2. Bed depth.
3. Ratio of air to fuel.
4. Superficial gas residence time; i.e., decrease in superficial gas velocity.
5. Carbon content of fly ash; i.e., less inert matter would be available to absorb energy.

6. Mesh size of feed; i.e., decrease in feed particle size.

On the other hand, the cost of a fluidized-bed boiler might be favored by a low value of each of these parameters. Thus, it might be necessary to find an optimum combination for combustion efficiency, auxiliary power requirements and capital costs. The goal of the experimental design was to provide, with a minimum number of experiments in the FBC, a mathematical relationship between the control variables and the principal response, combustion efficiency. The mathematical relationship would show if (and how) it might be possible to optimize the system.

Combustion efficiency (CE) is defined as:

$$CE = (1 - \frac{H_o}{H_i}) \times 100 \qquad (5)$$

where $H_o$ = higher heating value of unburned fuel components leaving the system, Btu/lb x lbs.

$H_i$ = higher heating value of fuel entering system, Btu/lb x lbs.

### 6.2.2 Identification of Variables

As part of this program an attempt was made to identify the variables describing the fly ash combustion process in a fluidized-bed. The system was broken down into six major components: (1) the air feed; (2) the fuel feed; (3) the bed, (4) gaseous products, (5) gas-borne solids, and (6) the reactor. Three categories of variables were enumerated for each of the six components: independent (or control) variables; dependent (or response) variables; and design variables. The potentially important parameters were listed under each category. These are given below in Tables 1 through 6.

It should be recognized that the classification is somewhat arbitrary and that all controls influence all responses. A fourth category, natural material variables, was also established to show how the fuel might be characterized.

Some of the symbols may not be understood by all readers. It is unimportant since they are not used elsewhere in the report.

TABLE 1. COMBUSTION AIR FEED VARIABLES

---

A. Control Variables

1. Mass flow rate, lb/hr $ft^2$
2. Temperature above ambient, $^{\circ}F$
3. Pressure, atmospheres
4. Composition; e.g., % oxygen, % humidity

B. Design Variables

1. Distributor design
2. Injection point of secondary air
3. Arrangement of secondary injection system

C. Response Variables

None

## TABLE 2. FUEL FEED VARIABLES

---

A. Control Variables

1. Mass flow rate, lb/hr $ft^2$
2. Temperature above ambient, °F
3. Composition; e.g., % carbon, % inert diluent, % moisture
4. Particle size distribution
5. Injection gas rate; lbs of solid /lb of injection gas
6. Use of homogenous combustion catalyst

B. Design Variables

1. Injector design
2. Injector location
3. Recycle rate
4. Injection gas selection; e.g., air, compressed flue gas, steam

C. Response Variables

None

D. Natural Material Variables

1. Ignition temperature
2. Porosity
3. Ash particle size
4. Ash sintering temperature
5. Degree of graphitization of carbon
6. Ash/carbon matrix characteristics
7. Attrition characteristics
8. Swelling characteristics
9. Surface area
10. Shape factor
11. Density
12. Emissivity
13. Diffusivity

## TABLE 3. BED VARIABLES

---

A. Control Variables

1. Mass, pounds (weight)/square foot of grid area
2. Composition; e.g., particle size distribution, particle density, bulk density, thermal properties, hardness
3. Temperature history; e.g., is the steady-state temperature approached from a higher or lower initial temperature

B. Design Variables

1. Combustor shape; e.g., constant cross section or conical
2. Internals; e.g., arrangement of heat transfer surfaces, inclusion of a knockout; i.e., baffles in freeboard
3. Circulation of hot or cold particles
4. Heat sink; e.g., surface area, effectiveness, temperature, location
5. Solids removal method
6. Cross-sectional area, $ft^2$

C. Response Variables

1. Nominal or average bed temperature, $^oF$
2. Bed temperature distribution; e.g., $T(x,y,z)$
3. Effective thermal conductivity - axial and horizontal, $Btu/ft^2hr^oF/ft$
4. Diffusivity - axial and horizontal, $ft^2/hr$
5. Expanded bed height, ft
6. Bed density distribution; e.g., $\rho\ (z)$
7. Bed circulation patterns, bubble frequency, channel life and movement
8. Equilibrium particle characteristics; e.g., particle size and shape
9. Residence time distribution of fine particles, $t_r(d_p)$
10. Fines capture rate through agglomeration
11. Ash particle size increase through agglomeration
12. Viscosity of fluidized-bed

---

## TABLE 4. COMBUSTION-GAS STREAM

A. Control Variables

None

B. Design Variables

1. Arrangement of above-bed heat transfer surfaces
2. Volume of freeboard
3. Laminar or turbulent flow

C. Response Variables

1. Mass flow rate, lb/hr (or superficial velocity, ft/sec)
2. Final composition; e.g. % $O_2$, $N_2$, $CO_2$, CO, $H_xC_y$
3. Composition distribution; e.g., $O_2(x,y,z)$ $SO_2(x,y,z)$
4. Temperature distribution
5. Final or average temperature, $^oF$
6. Solids loading, grains/scf
7. Electrical properties of gas (for electrostatic precipitation)

## TABLE 5. FLUE GAS SOLIDS VARIABLES

A. Control Variables

None

B. Design Variables

1. Freeboard (disengaging height)
2. Above-bed baffles
3. Secondary air injection

C. Response Variables

1. Mass flow rate, lb/hr
2. Final composition, %C, $\%H_2$, %S, $\%N_2$, %Cl, %NA+K
3. Size and distribution and density distribution
4. Resistivity (for electrostatic precipitation)

TABLE 6. REACTOR VARIABLES

---

A. Control Variables

1. Position of cooling surfaces; e.g., in-bed, above bed
2. Coolant temperatures

Note: There were also included in Table 3 as design variables. Their inclusion here is to emphasize the role of heat removal as a control variable in an experimental apparatus

B. Design Variables

(Covered under fuel, air, bed, and products of combustion)

C. Response Variables

1. Steam generated or superheated; i.e., energy removed from bed via heat exchange surface
2. Erosion of surfaces
3. Corrosion of surfaces
4. Slagging of surfaces

---

While the listing in Tables 1-6 is extensive, it cannot include all of the variables which actually influence the performance of a combustor nor all of the variables by which this performance might be measured.

### 6.2.3 Selection of Most Important Variables

Of the variables listed in Tables 1 through 6 those considered most significant are listed below:

As important control variables the following would be included: Mass flow rate of air (1.A.1), Mass flow rate of fuel (2.A.1), Composition of fuel (2.A.3), Particle size distribution of fuel (2.A.4), Mass of bed (3.A.1), Position of cooling surfaces (6.A.1) and Coolant temperatures (6.A.2).

As important response variables would be included: Nominal or average bed temperature (3.C.1), Superficial velocity (4.C.1), Final gas composition (4.C.2), Final gas-borne solids composition (5.C.2) and Energy removed from bed heat exchange surfaces (6.C.1).

If all of these most important control and response variables had been studied thoroughly, it might have been possible to achieve a relatively basic understanding of the process of fluidized-bed combustion of carbon in fly ash and fluidized-bed heat exchange. Unfortunately it did not appear possible to conduct such thorough fundamental studies with the financial resources which were available. On the other hand a basic understanding of the process was not required to (a) prove that it would work and (b) produce a prediction tool for use in design.

Even these limited goals could only be partially achieved: (a) Definitive proof or operability must await the construction of a large system, (b) No means were available at the outset of the program with which bed temperature could be predicted and no such technique was produced as a result of the program described here. Instead of being a response variable as anticipated beforehand, bed temperature had to be treated as a control variable. This anomalous behavior of bed temperature will be discussed later.

### 6.2.4 Variables Actually Studied

In preparing the proposed work plan in August, 1969, it was useful to treat the variables somewhat differently than indicated in the preceding paragraphs. Table 7 below lists the variables to be studied just as they were listed in the proposed work plan, the range originally suggested and the range actually studied.

It is instructive to use this original format and to explain how response variables, as listed in Tables 1 through 6, can appear as control, or, possibly a better term, pseudo-control variables.

TABLE 7

CONTROL AND PSEUDO-CONTROL VARIABLES SELECTED FOR CARBON-BURNUP CELL INVESTIGATION

| | Variable | Range Initially Suggested | Range Actually Studied |
|---|---|---|---|
| 1. | Carbon content of fly ash | 20% to 80% | 28% to 65% |
| 2. | Particle size of fuel | as received and 70% through 200 mesh | as received |
| 3. | Static bed depth | 6" to 18" | 10" to 22" |
| 4. | Nominal bed temperature | 1500°F to 2100°F | 1700°F to 2150°F |
| 5. | Excess air | 0% to 50% | 10% to 90% |
| 6. | Superficial velocity | 5 to 20 fps | 6 to 15.3 fps |

Temperature could be treated as a control variable since it was possible to alter the bed temperature during the course of an experiment by inserting or withdrawing a heat sink in the form of a bayonet tube. True, the temperature changed internally in response to an external control, the position of the bayonet, but not in a way that could be predicted. The tube was moved up and down until the desired temperature was achieved.

Excess air (actually residual oxygen), could more readily be used as the control than air rate. Residual oxygen was indicated and recorded at the FBC operating station. whereas air rate was indicated via a manometer which required interpretation.

Superficial velocity was treated as a control variable in planning since the selection of this parameter determined the sizing of the starter bed.

The other control variables listed in Table 7 -- carbon content of fly ash, particle size of this fuel and the static bed depth -- were also not subject to absolute control. The fuel would vary from test to test and quite probably within tests since it was real fly ash; not even homogenous, let alone "reagent grade". Static bed depth was as much a response as a control. The bed could increase in depth via ash agglomeration or decrease due to attrition. Tests were never of sufficient duration for the bed depth change to become important.

The statistical consultant, Professor Arthur E. Hoerl of the University of Delaware, selected from among the various controls, pseudo-controls and methods of expressing the values the following:

Bed temperature, static bed depth, carbon feed rate, inert feed rate and air feed rate. In addition to these control variables, it was found desirable to add some special experiments which might point up design considerations. These special tests included the addition of coal and operation with two bed particle-size distributions. The effect of coal was tested since in an actual Carbon-Burnup Cell coal will be used to maintain a constant bed temperature despite variations in flyash flow and fuel value.

The effect of coal addition and of bed particle size could not be included in the modeling process. Neither was found to be of major importance over the range studied.

The effect of grinding the fly ash was omitted from the test program since favorable results were obtained without this step.

### 6.2.5 Preliminary Modeling (A Simple Physical Model)

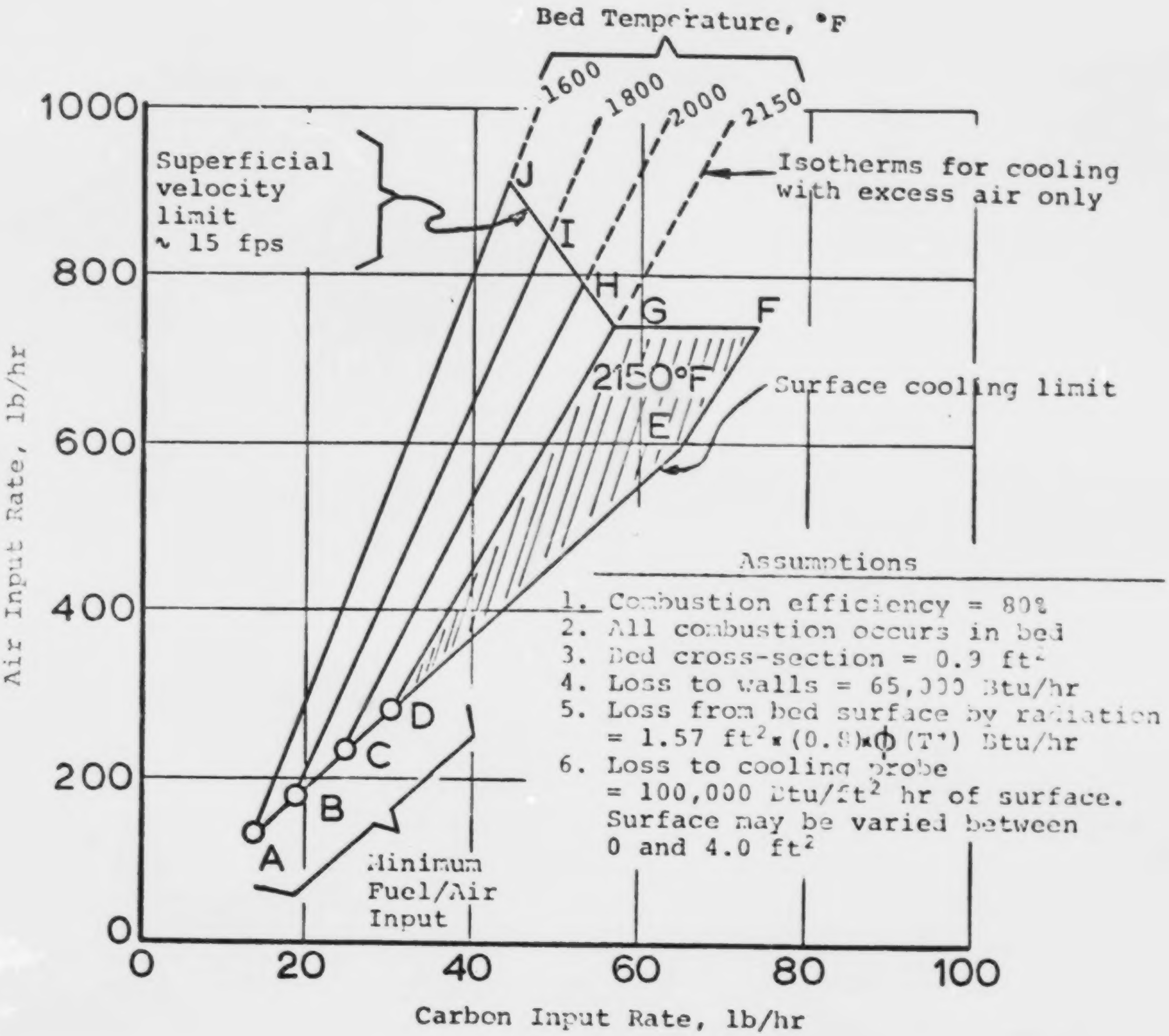
Before the experimental work began, a preliminary physical model was prepared in order to aid in the statistical design of the experiments. The model was intended to serve only as a map of the operating region of the FBC. This model is shown diagramatically in Figure 16.

It was assumed for the purposes of this model that the heat release rate within the fluidized-bed was given by multiplying together the carbon feed rate, the heat of reaction for

$$C + O_2 \rightarrow CO_2$$

and an arbitrary combustion efficiency, taken as 80%.

The heat loss rate from the bed through contact with walls of the combustor was taken as 65,000 Btu/hr; a heat loss through radiation was taken as proportional to the fourth power of the bed temperature with emissivity-absorptivity factors taken as 0.8 from a radiating surface of 1.57 square feet; a sensible heat loss in the flue gas was assumed to be proportional to the bed temperature, the air rate and the flue gas $CO_2$ content; and heat removed by the variable cooling coil, was taken to be the product of the heat flux (100,000 Btu/hr ft$^2$



Assumptions

1. Combustion efficiency = 80%
2. All combustion occurs in bed
3. Bed cross-section = 0.9 $ft^2$
4. Loss to walls = 65,000 Btu/hr
5. Loss from bed surface by radiation = 1.57 $ft^2 \times (0.8) \times \phi (T^4)$ Btu/hr
6. Loss to cooling probe = 100,000 Btu/$ft^2$ hr of surface. Surface may be varied between 0 and 4.0 $ft^2$

**FIGURE 16** PREDICTED AIR-CARBON OPERATING REGIME OF THE FBC

of surface at 2150°F) times the exposed coil surface (varied between 0 and 4.0 $ft^2$). In this model bed temperature can be reduced either by increasing excess air, inserting the cooling coil or both.

With the equations thus derived, it follows that the bed temperature would be predicted in this apparatus by the control variables; carbon rate, air rate and exposed coil surface. The baseline of Figure 16, line A-D, is for stoichiometric carbon/air feed. No excess air and no cooling, other than the wall and radiation, is provided to control bed temperature. The bed temperature is shown to increase from point A (1600°F) to point B (1800°F) as the carbon and air rates are simultaneously increased in stoichiometric proportions. Further increase in both fuel and air, maintaining a constant and stoichiometric fuel/air ratio, is shown to increase the bed temperature to 2000°F at point C and 2150°F to point D.

In preparing the map, 2150°F was considered maximum feasible operating temperature since the bed could become sticky and collapse if the temperature were increased further. However, operation at 2240°F had been demonstrated in one experiment with coal. To limit the temperature to 2150°F as the fuel and air rates are increased beyond point D toward point E, additional cooling must be provided. This additional cooling is provided by the insertion of the cooling coil. At point D, 0 $ft^2$ of surface is exposed; at point E, the 4.0 $ft^2$ of the coil are fully exposed to the bed.

A heat balance shows a stoichiometric fuel-air ratio is no longer possible yond point E, if the temperature is to be maintained at 2150°F, since the coil is fully inserted. Therefore an increase in fuel rate must be accompanied by more than a stoichiometric increase in air rate so as to provide cooling. Line E-F is constructed by calculating the excess air required to maintain a 2150°F bed temperature as the fuel rate is increased. At point F the temperature and air rate (approximately 750 lb/hr) provide a superficial gas velocity of 15 feet per second. This gas velocity was arbitrarily chosen as the maximum permissible. An increase in air rate beyond that at point F is not permitted until the bed temperature is reduced below 2150°F. Point F also represents the maximum permissible firing rate, equivalent to a heat release rate of 950,000 Btu/hr per square foot of grate area.

Line A-J is the 1600°F isotherm. Along this line air rate is increased more rapidly than the fuel rate so as to maintain a constant bed temperature without the use of the cooling coil; i.e., the excess air provides the required cooling. At point J the air rate (approximately 900 lb/hr, 40 lb/hr fuel) and the 1600°F bed temperature would give a superficial velocity of 15 ft/sec. Point J therefore represents the maximum permissible air rate. Isotherms similar to line A-J can be drawn at other temperatures (See lines B-I, C-H, D-G on Figure 16) for the case where the required cooling is provided by excess air.

Line J-G represents a constant velocity curve connecting those points on each of the isotherms where the air rate and bed temperature result in the 15 fps gas velocity set as the maximum. For example, at point I the fuel rate has been increased to 40 lbs/hr while the excess air was decreased sufficiently to maintain the superficial gas velocity at 15 fps while the bed temperature rose to 1800°F. The cooling coil is not in use in this area of the figure. Further decrease in excess air by increasing fuel and decreasing the air rate past point I maintains the 15 fps limit through point H ($2000^{o}$F) to point G ($2150^{o}$F), the maximum permissible temperature. Between points G and F the cooling coil must be inserted as the fuel rate is increased so that the bed temperature can be maintained at $2150^{o}$F. The superficial velocity remains 15 fps.

Within the area enclosed by lines A-E, E-F, F-G, G-J and J-A, the Carbon-Burnup Cell was assumed to be stable.

Assuming a combustion efficiency different from 80% would result in an area of similar shape. A three-dimensional body could then be constructed from the isoefficiency maps within which a Carbon-Burnup Cell would be assumed to be operable. At 100% combustion efficiency, the map would be displaced downward and to the left. At about 35% combustion efficiency, the regime of operation would appear as a single point at approximately 900 lb/hr air rate, 90 lb/hr-carbon and $1600^{o}$F.

Additional dimensions might be added to the regime of operation representing other variables; for example, the temperatures of the incoming air and fuel feel could be included in the physical model.

When the experimental work aimed at developing a statistical model was begun some of the test results actually fit the simple physical model depicted in Figure 16; most did not. As will be shown later it was not possible to predict the bed temperature as a function of fuel rate, air rate, combustion efficiency and heat removal from the combustion chamber.

Three reasons will be offered for the deficiencies in the physical model. In summary, these are as follows:

1) The process is bi-stable; i.e., two apparently identical experiments will yield two different results depending, for example, on whether the steady state bed temperature was approached from an initially higher or from an initially lower value.

2) The physical model assumed that all energy released by the burning fuel occurred within the dense phase of the fluidized bed. An unknown fraction of the burning actually occurs in the freeboard above the bed.

3) The mechanisms by which the turbulent fluidized-bed loses energy to heat transfer surfaces above the dense phase surface are not fully understood. Radiation from the refluxing bed particles is one important mechanism for losing energy from the bed.

Despite these deficiencies, no attempts were made to improve the physical model since it served its mapping purpose adequately. However, fundamental studies of heterogenous combustion within a fluidized-bed would improve the model and have been advocated by Pope, Evans and Robbins (19) as well as others (15, 20). Fundamental studies on the course of combustion in a fluidized-bed might also be useful for air pollution control (12).

### 6.2.6 The Design of the Experiments: Selection of Test Conditions

The simple physical model and the predictions based on previous experience were reviewed with the statistical consultant. Because of the numerous parameters the consultant suggested that it would be desirable to run well over 100 tests. This was not possible with the available resources.

A total of 55 tests were run, 38 of which were used in producing the statistically derived performance model. All of the tests are described in Appendix E; while those used by the consultant are listed in his report, included here as Appendix F.

At the consultant's request, tests were run in small blocks, and the data reduced and the results reviewed following each block. The consultant would then select the conditions for the next small block of tests.

The design of the experiments was thereby conducted intermittently during the test program and did not actually precede it as the term design implies.

One of the first tasks assigned by the consultant was to explore the boundaries of the Carbon-Burnup Cell process; i.e., the regime of stable operation. This topic is discussed in the next section.

## 6.3 General Results: Exploring the Regime of Stable Operation

The physical model of Figure 16, and its extensions to a third dimension, combustion efficiency, indicated the outermost limits in which the Carbon-Burnup Cell system would be operable.

It should be recalled that the FBC only simulates a Carbon-Burnup Cell and problems encountered in the operation of this small system might not exist in a much larger device; the destruction of the uncooled metal walls is an obvious example. Conversely, problems in fuel distribution do not exist in a small bench scale system but must be considered in the commercial design.

An experimental tool, such as the FBC, is capable of answering questions about a process if these questions are properly posed. It is often impossible to determine if the apparatus has answered a question about its own limitations or about a limitation inherent in the process itself. One obvious example of this kind of ambiguity are the upper and lower temperature limits at which the process was stable. This and several other topics are discussed below.

### A. Low Temperature

It was found that operations below about 1800°F were not feasible with fly ash as the fuel. The minimum temperature 1600°F used in the preliminary model had been chosen as a lower limit since coal combustion was stable at 1600°F and, more important, because a high ash ( ≈ 50%) char had been burned at 1600°F. Operation at low air rates (say, equivalent to a superficial velocity of 0.5 fps) should have been possible even at 1300°F in an almost adiabatic apparatus; this is impractical however. Only a few tests were conducted at 1800°F and below.

### B. Superficial Velocity

It was known, and confirmed in the tests, that a fluidized bed of coarse particles has a relatively small value for the Operable Velocity Ratio:

where

$$\text{Operable Velocity Ratio} = \frac{V_{max}}{V_{min}}$$

where

$V_{max}$ = Superficial velocity at which substantial elutriation occurs

$V_{min}$ = Superficial velocity at which substantial stagnation occurs

For particles in the size ranges -8 +16 and -16 +22, this ratio is on the order of 2.

A precise value for the ratio depends on the reactor configuration (e.g., higher values would be found for conical reactors), the operating temperature, and the nature of the particles. This last factor, the nature of the particles, relates to the definition of substantial stagnation. It is defined here as being that point of gas velocity and bed temperature at which the particles not only stagnate but also sinter into a large mass. Goldberger (17) has provided a curve that indicates, for his apparatus, a relationship between superficial gas velocity and bed collapse temperature.

### C. High Temperature

Sintering of the particles was found to occur at a bed temperature above 2000°F at the higher superficial velocities. This sintering was reflected by an increase in the particle size as fine matter released by the burning fuel adhered to the bed particles. The rate of growth appeared very rapid above 2100°F.

Agglomeration can be a potentially useful effect; it has been proposed that this particle capture effect be utilized to reduce the dust load to the pollution control system. However, it was not desirable here. Agglomeration in the FBC would require that either a system be provided to remove particles continuously, with separation of the larger fraction and return of the finer fraction, or that the air rate be increased continuously as the test progressed and the particles grow.

The former action was beyond the scope of the current effort. Continuous increase in air rate would have been useful if the apparatus had been instrumented so that the results could have been properly interpreted. Studies of both size maintenance and dynamic testing would appear to be justified, on the basis that the first may eventually be required in a commercial fluidized-bed boiler, and the latter is required if substantive insights on the nature of fluidized-bed combustion are to be achieved.

### D. Bed Particle Size vs. Temperature and Air Rates

In the tests performed, the operating regime of the Carbon-Burnup Cell system was limited to a lower temperature of 1800°F, an upper limit of 2050°F, and two bed particle size regimes. Air rates in the range below 800 lb/hr could be accommodated with a bed of particles sized -16 +22 U.S. Standard sieve size. Air rates above 600 lb/hr could be accommodated with a bed of particles sized -8 +16 U. S. Operation at 600 lb/hr was possible with either bed so that the effect of particle size on combustion efficiency could be determined.

### E. Bed Depth

The effect of bed depth was also explored. At the outset of the program, it was thought that operation with beds as shallow as 6" could be useful since coal had been burned in 4" deep bed. It was found, however, that below about 14", combustion efficiency deteriorated badly at the higher air rates.

The dependence of combustion stability on a minimum bed depth was assessed by a theoretical analysis of the rate of particle heating. It was assumed that initially the particles of fly ash would be at the temperature of the injection air. Further, the particle temperature rise would be equivalent to the air temperature rise. This was supported by an analysis of the rate at which energy is transferred to the particle, and the rate at which energy received by the surface of the particle would lead to a temperature rise in the center of the particle. For particles of carbon, less than 200 microns in diameter, the convective heat transfer rate to the particle and the rate of internal heat transfer were sufficiently high to insure that the particles and gas would be at essentially the same temperature during the heating-up period. The problem was reduced, therefore, to one of predicting the rate at which the gas temperature would rise as it entered the hot fluidized-bed. The analysis, presented as Appendix G, shows that within about 1/2" of the injection the fly ash particles should be close to the bed temperature.

One possible reason for the poor performance of a shallow bed might have been inadequate distribution of the fly ash by the fuel injector. The FBC feeder was not modified. Instead, beds of 10" and above were studied. The CBC fly ash feeder was changed to the mushroom arrangement as noted in the description of the equipment.

F. Other Limitations

There were a number of other limitations which may relate to the process but are probably related to the apparatus. For example, it was not possible to fire fly ash with a carbon content below about 30%. Fly ash with a high moisture content performed poorly. The overall system heat balance did not permit operation at above 10% residual oxygen in the flue gas (high excess air). The heat removal rate was also limited by heat balance considerations. When the ratio (heat removal from the chambers via the walls, hood and cooling probes plus dust) : (heat released) exceeded about 35% to 45% the combustion process would begin to deteriorate. It is believed that the fire would go out if this ratio is > 50%.

## 6.4 Statistically Derived Model for Fly Ash Carbon Combustion in a Fluidized-Bed Combustor

### 6.4.1 Description of Model

Based on the data from 38 (Appendix F) of 55 (See Appendix E) combustion tests in the FBC, an empirical regression model was developed with which the combustion efficiency of fly ash in the FBC may be accurately predicted if the following are known: (a) the bed temperature, (b) the air rate, (c) the static bed depth, (d) the fuel rate and (e) the carbon content of the fly ash.

Of the 17 tests not used in the model development, about one-half were performed after the model was complete, and one-half were too deviant to use or were conducted with coal as well as fly ash feed.

The general function is as follows:

$$y = b_0 + b_1 (T) + b_2 (A) + b_3 (H) + b_4 (C) + b_5 (I) + b_6 (I \times C) \tag{6}$$

where

$y$ = a particular response

$b_0$ to $b_6$ = constants determined in the analysis for the particular response

and

$T$ = bed temperature, $^{o}F$

$A$ = air rate, lbs/hr

$H$ = height of static bed, inches

$C$ = carbon rate, lbs/hr

$I$ = inert rate, lbs/hr

$(I \times C)$ = cross product of carbon rate and inert rate, $lbs^2/hr^2$

The parameters C and I are related to the fuel feed rate by the expressions

$$C = F \times (c) \tag{7}$$

$$I = F - C = F \times (1-c) \tag{8}$$

where

$F$ = fuel feed rate, lbs/hr

$c$ = carbon content of fuel, lb carbon/lb fuel

It has been noted previously that the bed temperature, T, appearing in Equation 6 as an independent variable, would, preferably have been treated as a response variable. However, temperature could not be treated as a response variable. When the heat removal rate, expressed in Btu/hr, by the cooling surfaces and in dust (the preferred independent variable) was employed in Equation 6, replacing the temperature, and when a correlation was then attempted to describe T as the response variable (y), no correlation was possible. This lack of correlation between T and the independent variables is attributed to the hypothesized bivariance of temperature (see Sec. 6.5.3).

The use of bed temperature as an independent variable in the performance model is acceptable since, in the commercial design, the bed temperature and its rate of change, will be used to vary the supplementary coal feed rate, or the preheat of CBC air.

The important response variables used for modelling were: the combustion efficiency, CE, and the residual oxygen, RO. The other gas constituents which were treated as response variables were $CO_2$, $SO_2$ and NO.

### 6.4.2 Results

The results of the tests, as described in Appendix F, were utilized to:

(a) determine the form of the model which best related all of the data (Equation 6 was the result, and

(b) compute the regression coefficients, $b_0$ through $b_6$, for the models. The values in Table F-2 (Appendix F, page F-9) were the result. The coefficients for CE, RO and $CO_2$ are also listed in Table 9.

In developing the form of the model, the principal response, CE-combustion efficiency, served as the primary criterion for evaluating alternative forms.

The reliability of the models expressed in terms of their respective F-ratios and $R^2$ is given below in Table 8.

TABLE 8

RELIABILITY OF CARBON-BURNUP CELL
PERFORMANCE MODEL (EQUATION 6)

| RESPONSE, Y | F-RATIO* | $R^2$** |
|---|---|---|
| CE (Combustion Efficiency) | 74.5 | .935 |
| CO ($O_2$) | 26.3 | .836 |
| $CO_2$ | 23.0 | .816 |
| $SO_2$ | 9.2 | .642 |
| NO | 6.0 | .535 |

*The F-ratio is a statistical measure of the variance in the response data which is explained by the model divided by the residue (unexplained) variance. If there were no relationship between the independent variables and the response, the F-ratio would tend to take on a value of one. In general, the higher the F-ratio, the more reliable the model is in explaining the trends in the data.

**The parameter $R^2$ is defined as the explained variation divided by the total observed variation. As $R^2$ approaches unity the model becomes more reliable.

---

It may be seen from Table 8 that the models for CE, RO and $CO_2$ as responses are reliable while those for $SO_2$ and NO are not.

Another method of demonstrating the reliability of a model is to plot the observed results against the calculated results. Figures 17 and 18 present such plots for CE and RO

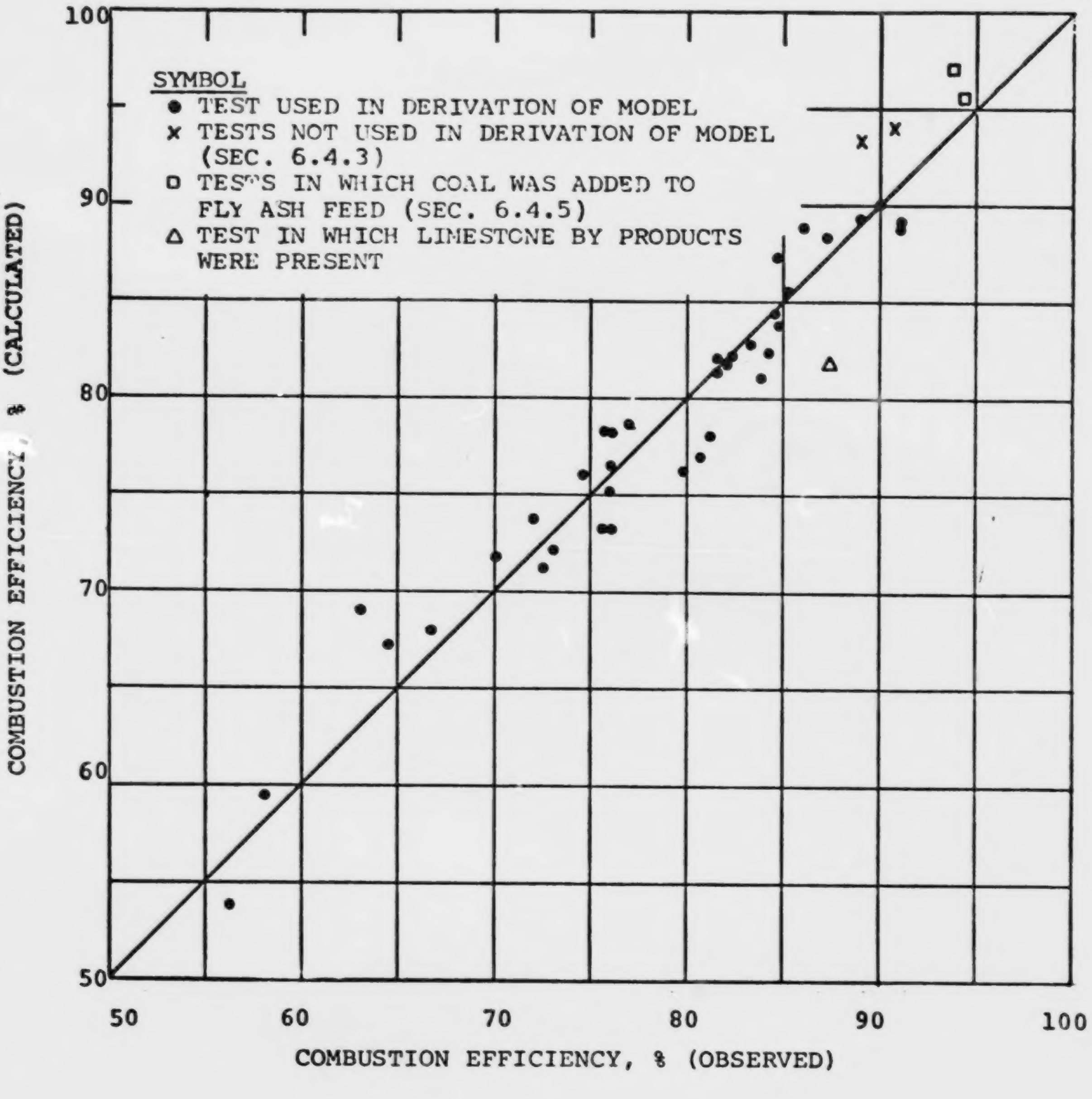


FIGURE 17 COMPARISON OF OBSERVED AND CALCULATED COMBUSTION EFFICIENCY IN FBC TESTS

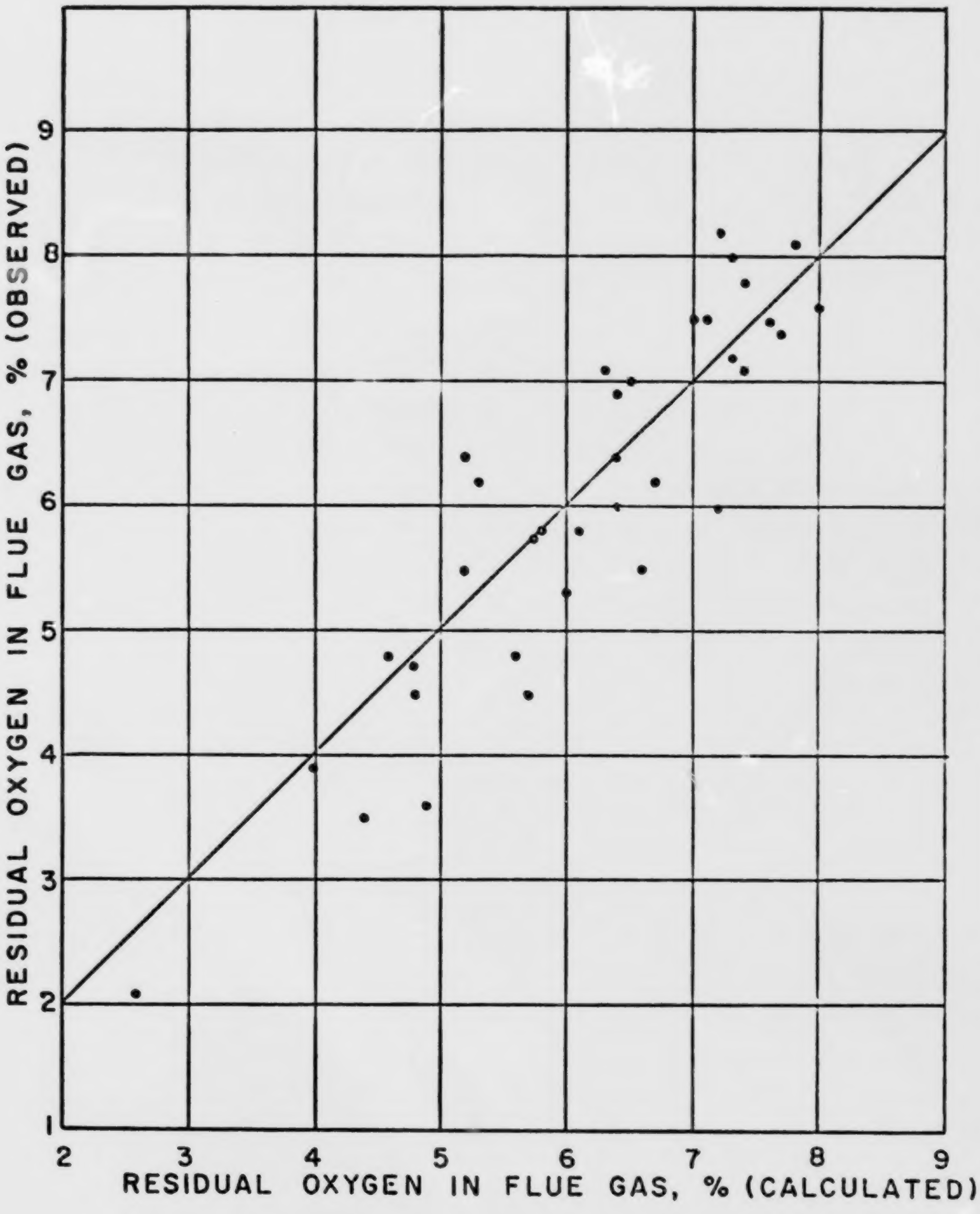


FIGURE 18 COMPARISON OF OBSERVED AND CALCULATED RESIDUAL OXYGEN LEVEL

respectively. The $45^{o}$ line represents an ideal correlation.

The regression coefficients for CE, RO and $CO_2$ listed in Table 9 will be used in sample calculations which follow immediately. ($SO_2$ and NO are discussed later in this section).

### 6.4.3 Test of Model in Same Apparatus

To illustrate the use of the basic model equation, Equation 6, for predicting combustion efficiency, residual oxygen and carbon dioxide, the data for FBC tests C-319-1 and C-319-2 will be evaluated. FBC tests C-319-1 and C-319-2 were performed after the derivation of the model was complete and were not used in the model preparation. In addition, the apparatus had been rebuilt to reduce the heat loss to the enclosure. So that the meaning of each parameter will be clear, a sample calculation is provided on page D-1.

TABLE 10

ILLUSTRATION OF CARBON-BURNUP CELL PERFORMANCE MODEL (EQUATION 6)

| | Test No. | |
|---|---|---|
| | 319-1 | 319-2 |
| Combustion Efficiency, % | | |
| Calculated | 93.3 | 94.0 |
| Observed | 89.0 | 90.6 |
| Residual | - 4.3 | - 3.4 |
| Carbon Dioxide, % | | |
| Calculated | 13.5 | 12.9 |
| Observed | 11.8 | 11.8 |
| Residual | - 1.7 | - 1.1 |
| Residual Oxygen, % | | |
| Calculated | 7.3 | 7.9 |
| Observed | 8.2 | 8.7 |
| Residual | +0.9 | +0.8 |

TABLE 9. REGRESSION COEFFICIENTS FOR USE IN RESPONSE FUNCTION (EQUATION 6)

| TERM TO WHICH COEFFICIENT IS APPLIED AND APPLICABLE RANGE | DESIG-NATION | DIMENSIONS | COEFFICIENTS* | | |
|---|---|---|---|---|---|
| | | | For Combustion Efficiency | For Carbon Dioxide | For Residual Oxygen |
| Constant | $b_0$ | None | - 13.78 | - 2.498 | + 22.91 |
| Bed Temperature<br>(1,750 $\leq T \leq$ 2,140°F) | $b_1$ | $^{o}F^{-1}$ | + 0.05193 | + 0.007547 | - 0.007353 |
| Air Rate<br>(330 $\leq A \leq$ 820 lb/hr) | $b_1$ | hr/lb | + 0.0462 | - 0.01190 | + 0.01300 |
| Bed Height<br>(10 $\leq H \leq$ 22 in) | $b_3$ | $in^{-1}$ | + 0.3831 | + 0.1382 | - 0.1390 |
| Carbon Rate (C)<br>(26 $\leq C \leq$ 96 lb/hr) | $b_4$ | hr/lb | - 0.8737 | + 0.1589 | - 0.1769 |
| Inert Rate (I)<br>(31 $\leq I \leq$ 206 lbs/hr | $b_5$ | hr/lb | - 0.1905 | + 0.01259 | - 0.01756 |
| Cross Product (C) (I) | $b_6$ | $hr^2/lb^2$ | + 0.00270 | - 0.0002862 | + 0.0003587 |

*Coefficients are listed here as derived from FBC data. One method of applying the model to other apparatus is to assume that cross-sectional area is the appropriate scaling factor. This topic is discussed in Section 6.5.2.

The residuals for combustion efficiency are higher than those of the majority of tests used in preparing the model, but are not the most deviant (see Appendix F). This illustrates that the model is probably a useful engineering tool for predicting the performance of a Carbon-Burnup Cell from a knowledge of the ingredient rates, composition and the bed temperature.

A second test of the model was made when the results of the statistical analysis were reviewed. As presented in Appendix F there was a residual of +2.7% between the observed and calculated CE's for Test C-316-2. When the correct data were used the calculated value matched the observed value exactly.

### 6.4.4 Relative Significance of Parameters in Model Equation

In order to illustrate the relative importance of the control variables a simple parametric study was done using Equations 6, 7, and 8 and the regression coefficients in Table 9 for combustion efficiency.

#### 6.4.1.1 Effect of Bed Temperature, T

Table 9 shows that for each 19.3°F rise in bed temperature a 1% increase in combustion efficiency is predicted by Equation 6, all other independent variables remaining constant. Figure 19 shows the bed temperature at which a 90% combustion efficiency will be achieved as a function of energy release rate for several bed depths and with A/C = 14 (21% excess stoichiometric air), c = 0.5. One curve is shown for A/C = 17.3 (50% excess stoichiometric air). Solid lines are used where the result is reasonable, and

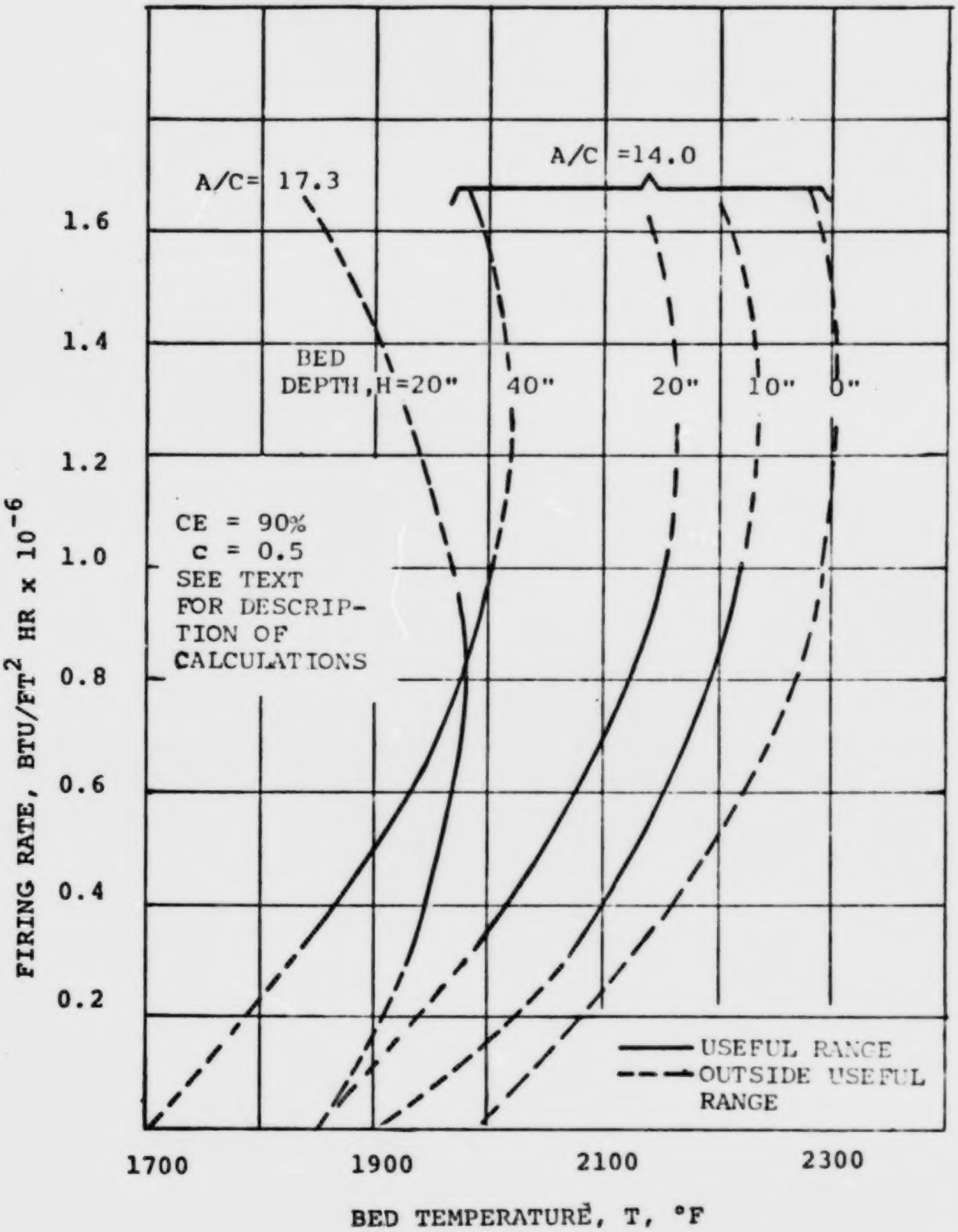


FIGURE 19 EFFECT OF BED TEMPERATURE ON CARBON-BURNUP CELL PERFORMANCE MODEL PREDICTIONS

dotted lines show Equation 6's predictions which are probably outside the equation's useful range. The curvature shown in Figure 19 is inherent in the Equation 6 because a square term enters as the cross-product of carbon rate and inert rate and this becomes significant at very high throughputs.

Figure 19 clearly illustrates that as the throughput is increased, the bed temperature (or bed depth) must be increased to maintain the desired 90% combustion efficiency. It is also seen that the air to carbon ratio may also be increased in lieu of bed depth or temperature to maintain the desired combustion efficiency.

### 6.4.4.2 Effect of Bed Depth, H

Table 9 shows that for each 3.6" increase in bed depth a 1% increase in combustion efficiency is predicted. Figure 20 shows the bed depth at which 80, 90 and 100% combustion efficiency will be predicted by Equation 6 as a function of throughput for a stoichiometric air-to-carbon ratio. The bed temperature will be maintained constant at 2000$^{o}$F and again, c = 0.5.

### 6.4.4.3 Effect of Air Rate, A

Table 9 shows that the combustion efficiency, predicted by Equation 6 for the FBC, will increase by 1% for each 22 lb/hr increase in air rate, equivalent to 25 lbs/hr ft$^2$ of grate area. This increase in air flow will also increase energy loss from the bed (in the form of sensible heat of the flue gases) by roughly 10,000 Btu/hr while the 1% increase in efficiency releases an additional 5000 Btu/hr. Therefore, to maintain a constant bed temperature, the heat transfer surface must be

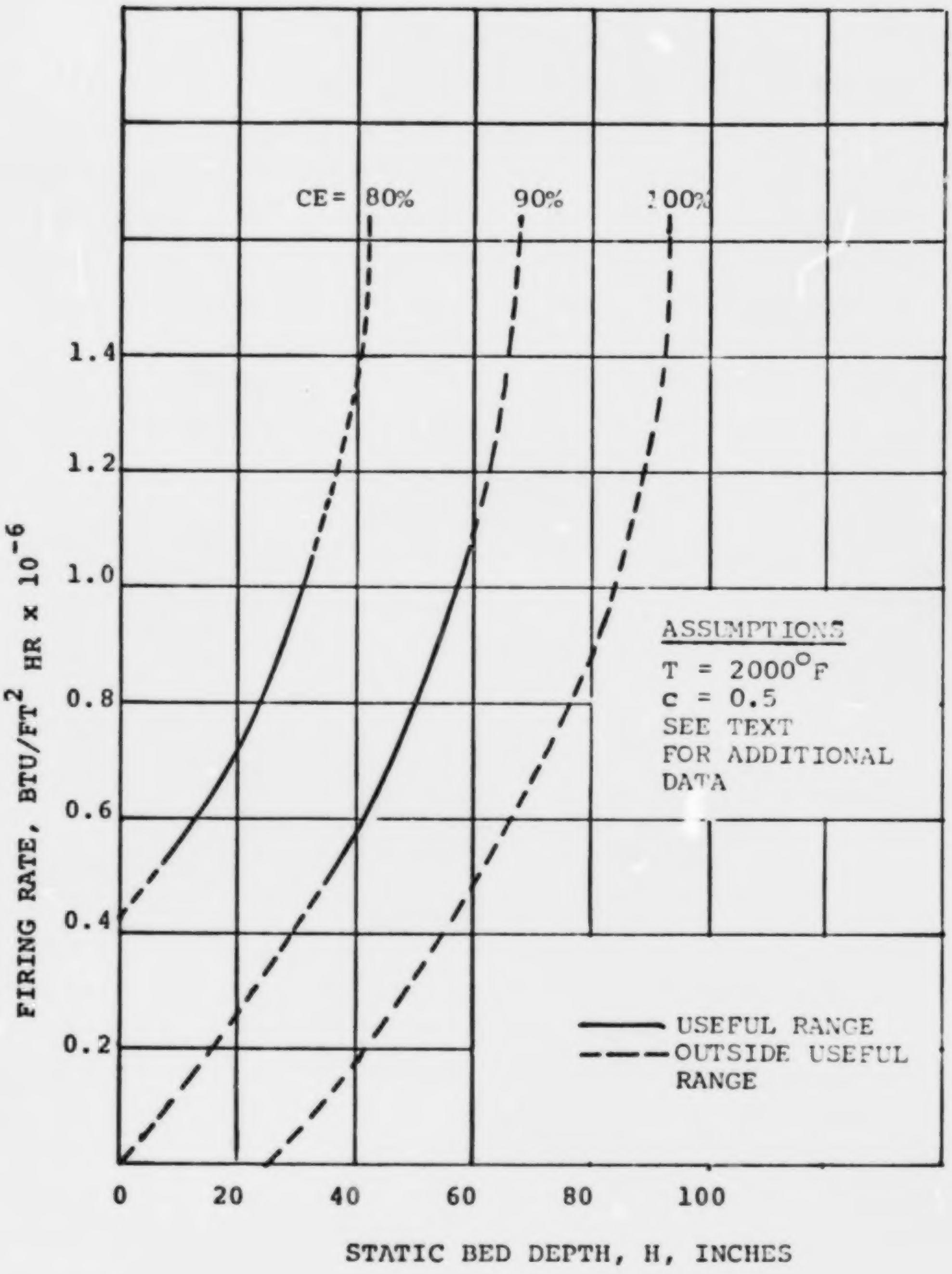


**FIGURE 20** EFFECT OF STATIC BED DEPTH ON CARBON BURNUP CELL PERFORMANCE MODEL PREDICTIONS

reduced by approximately 0.05 $ft^2$ (See Section 6.2.5).

A more useful means of describing the effect of air rate is to discuss the excess stoichiometric air as given by the ratio A/C. For $C + O_2 = CO_2$ the stoichiometric oxygen requirement is met with 11.53 pounds of air per pound of carbon. Figure 21 shows the excess stoichiometric air required for Equation 6 to predict 90% combustion efficiency as a function of throughput. Bed temperature is set at $2000^oF$, and bed depth at 24". The carbon content of the fuel is varied in this example.

### 6.4.4.4 Effect of Carbon Content of Fly Ash Fuel, c

To prepare Figure 21, Equation 5 was rewritten:

$$Y = b_o + b_1 (T) + b_2 (A/C)(C) = b_3 (H)$$

$$+ b_4 (C) + b_5 (I/C)(C) + b_6 (C)(I/C)(C) \qquad (6A)$$

From Equations 7 and 8 it may be seen that:

$$I/C = (\frac{1.0-c}{c}) \qquad (9)$$

where

$0 \leq c \leq 1.0$ = carbon content of fuel fly ash

Since "c" thus appears in the last two terms of Equation 6A it is not possible, as it was in the case of the previous examples for bed temperature and depth to state how a change in carbon content, c, will effect combustion efficiency without also selecting a value for the carbon rate, C. To illustrate the significance of the carbon content of the fly ash fed to a Carbon-Burnup Cell, Figure 22 was prepared. It is based on the following assumptions: C = 50 lbs/hr, the equivalent of a nominal firing rate of 820,000 Btu/hr $ft^2$, $T = 2000^oF$, H = 24" and A/C = 16.2. The pounds of carbon in the

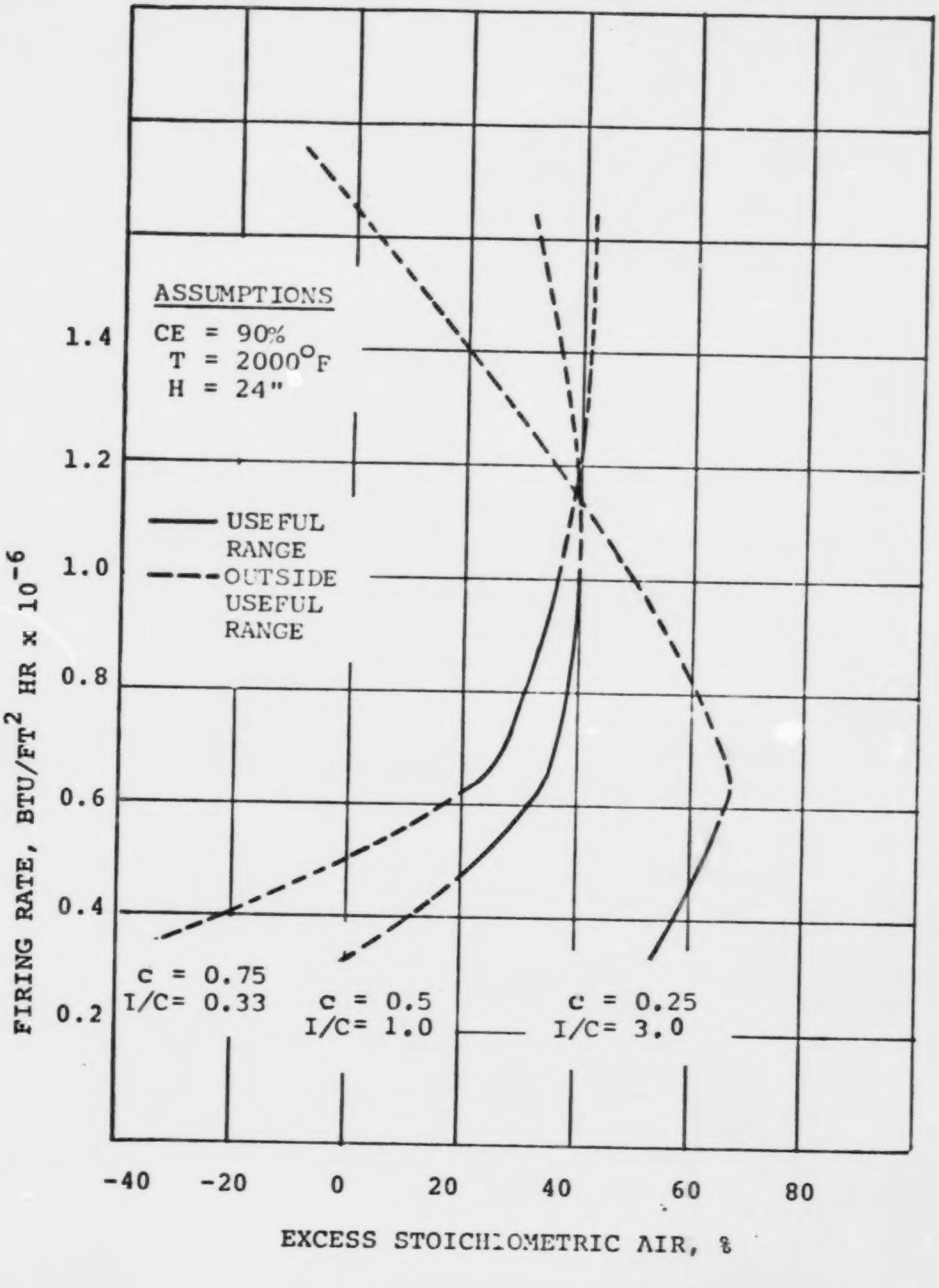
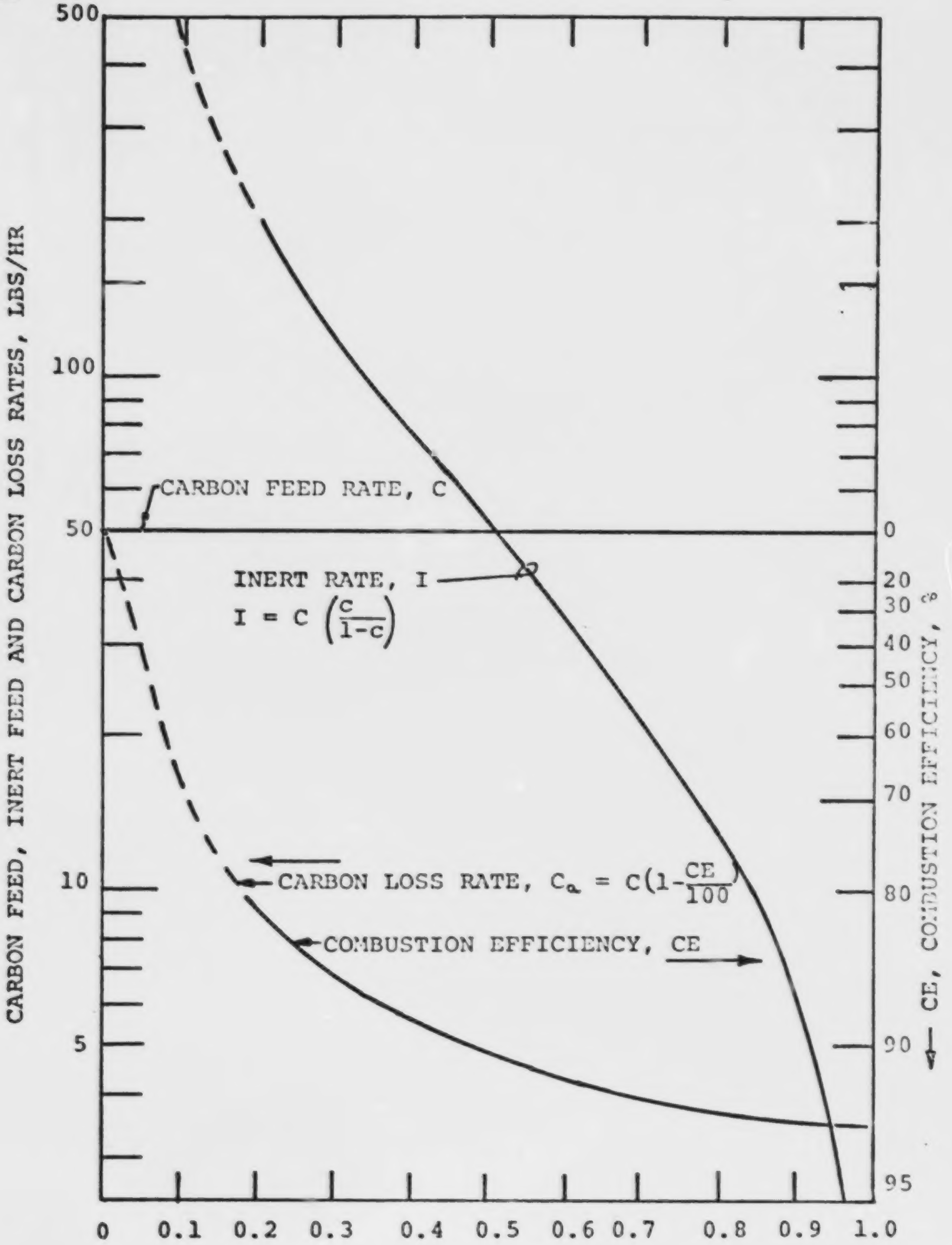


**FIGURE 21** EFFECT OF EXCESS STOICHIOMETRIC AIR ON CARBON-BURNUP CELL PERFORMANCE MODEL PREDICTIONS



c, CARBON CONTENT OF FLY ASH FEED, LBS C/LB FEED

FIGURE 22 PREDICTED EFFECT OF CARBON CONTENT OF FLY ASH FEED ON PERFORMANCE OF CARBON-BURNUP CELL

ash, $C_a$, predicted by the model equation and

$$C_a = C\left(1-\frac{CE}{100}\right) \tag{10}$$

is plotted as is the inert rate, I.

Figure 22 reveals that the smallest carbon loss is realized if the fuel is relatively rich in carbon; i.e., $c > 0.3$. The figure also reveals why a fuel low in carbon; i.e., $c \leq 0.2$ will not be a suitable feed to the Carbon-Burnup Cell of a Multicell Fluidized-Bed Boiler. As $c \rightarrow 0$ the inert rate becomes so high that it would be impossible to maintain combustion; the ash sensible heat would remove more energy than the fuel could supply. For the air rate and temperature chosen in the example, the amount of heat transfer surface becomes zero at $c \approx 0.2$. Below $c = 0.2$, a temperature of $2000^{\circ}F$ is no longer possible even in an adiabatic chamber.

The carbon content of the fly ash feed is not a true control variable in a Multicell Fluidized-Bed Boiler since the quantity and fuel value of the fly ash fed to the Carbon-Burnup Cell depends on the operation of the primary cells.

Although the topic of design will be discussed later, it will be useful to show here, by way of a sample calculation, how the combustion efficiency of the primary cells affects the operation of the Carbon-Burnup Cell by affecting the carbon content of the fly ash feed.

Assume that a coal having the following ultimate analysis is to be fed to a Multicell Fluidized-Bed Boiler:

$C = 72.0\%$, $H = 5.5\%$, $N = 1.5\%$, $O = 10.0\%$

$S = 3.0\%$, Ash $= 8.0\%$ where the values are in weight % for the elements shown. In this example, the hydrogen and oxygen values have been adjusted to include the moisture.

This is a deviation from the normal analysis format. Using Dulong's formula

$$\text{Btu/lb} = 145.44\text{C} + 620.28\ (\text{H} - \frac{\text{O}}{8}) + 40.50\ \text{S} \qquad (11)$$

the higher heating value of this coal is 13,229 Btu/lb. Assume that all of the mineral matter in the coal appears as fly ash and that 99.5% of the fly ash is captured in an electrostatic precipitator. This fly ash will be fed to the Carbon-Burnup Cell(s). Using the methods of the ASME Power Test Code (21) and the model equations, Figure 23 shows the combustion efficiency for various values of "combustible content of refuse"; i.e., loss-on-ignition of fly ash. Two values are shown for combustion efficiency in the primary cells: Curve "A" for the combustion efficiency of all the fuel constitutents of the coal C, H, S; and Curve "B" for the combustion efficiency of the carbon content of the coal alone. Note the loss-on-ignition of fly ash is assumed to be almost entirely carbon. Curve "C" shows the combustion efficiency for fly ash carbon fed to the Carbon-Burnup Cell, obtained from Figure 22). Curve "D" shows the overall combustion efficiency assuming that the Carbon-Burnup Cell fires the fly ash under the conditions used to generate Figure 22.

From Figure 23 it may be seen that the overall combustion efficiency of a Multicell Fluidized-Bed Boiler is relatively insensitive to the carbon content of the fly ash from the primary cells. This follows from the fact that as the primary cells become less efficient (Curves A and B) the Burnup-Cell can become more efficient (Curve D). Optimization is discussed in Section 8.

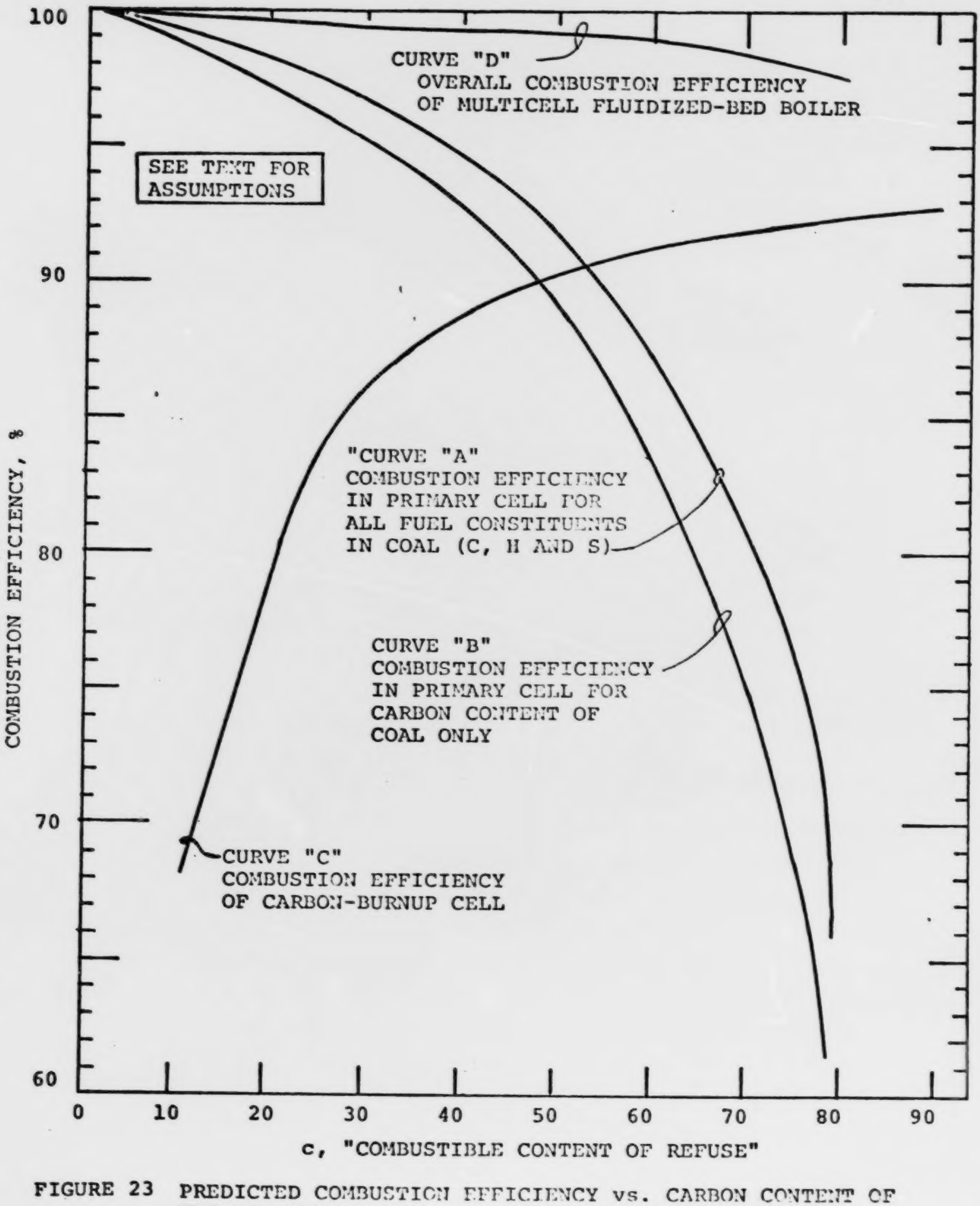


FIGURE 23 PREDICTED COMBUSTION EFFICIENCY vs. CARBON CONTENT OF FLY ASH FROM PRIMARY FLUIDIZED-BED BOILER CELL

### 6.4.4.5 Effect of Firing Rate, C

In Equation 6A, the parameter C appears in four terms. To illustrate the significance of this parameter, Figure 24 was prepared which plots as "Y" the combustion area per $10^6$ Btu of heat release, and as "X" the firing rate. X and Y here are defined by:

$$X = \frac{C,\ \text{lb/hr}}{0.86\ \text{ft}^2} \times 14{,}100\ \text{Btu/lb} = 16{,}400C,\ \text{Btu (fired)/ft}^2\ \text{hr}$$

$$Y = 60.99\ C\ \left(\frac{CE}{100}\right),\ \text{ft}^2/10^6\ \text{Btu (released) per hr}$$

For this example I/C = 1.0, H = 24" and T = 2000°F. Two values for A/C are used; A/C = 16.2 (40% excess stoichiometric air) and A/C = 11.6 (0% excess stoichiometric air).

It may be seen that the reduction in combustion efficiency which results from increasing throughput has little effect on the required unit size; the over-riding factor is throughput. Halving the firing rate ($\text{Btu/ft}^2\text{hr}$) would require doubling the area of the Carbon-Burnup Cell. It was not meaningful to plot combustion efficiency on Figure 24, instead 100-CE was plotted. It may be seen that combustion efficiency does decrease with increasing throughput.

### 6.4.4.6 Conclusion Regarding Significance of Parameters

The preceding paragraphs have discussed the effect of bed temperature, bed depth, excess air, firing rate and carbon content on the combustion efficiency of the Carbon-Burnup Cell as this efficiency is predicted by the model equation. It was noted that carbon content is not a true control variable although it is a design consideration. Each of the following changes will increase the combustion efficiency by 1%, all other

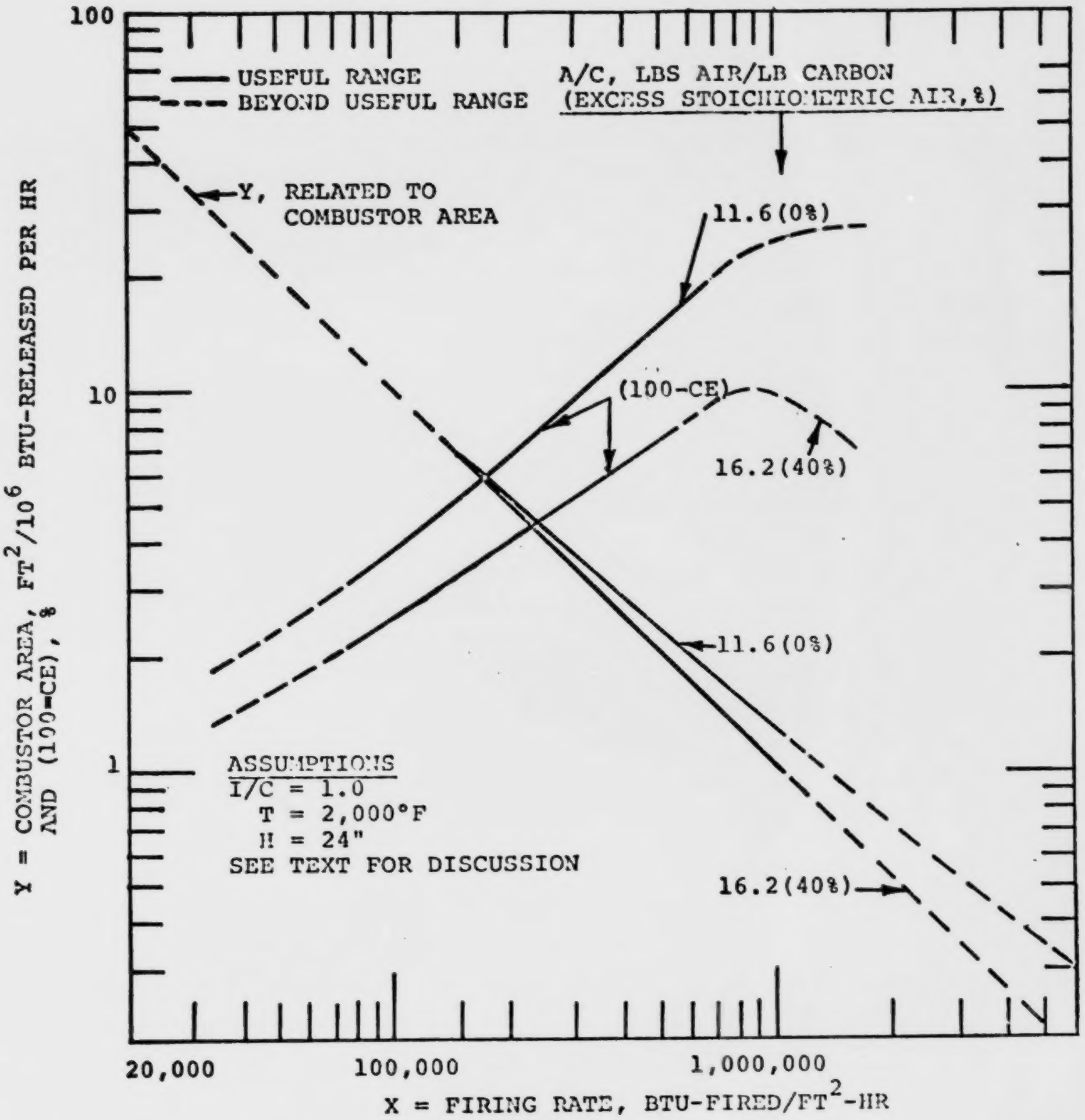


FIGURE 24 EFFECT OF FIRING RATE ON PREDICTED PLAN AREA OF CARBON-BURNUP CELL

parameters held constant:

a. An increase in air rate of 25 lbs/hr $ft^2$

b. An increase in bed depth of 2.6 inches

c. An increase in bed temperature of 19.3°F

As noted previously, each of these changes, a-c, may have an undesirable effect on either capital costs, operating costs or both. It should also be clear that in a real system any change in one parameter will affect all other parameters. Still, viewed in isolation, the following conclusions may be drawn:

a. The air rate (or really air to carbon ratio) would be increased until the gain in combustion efficiency is matched by the loss as sensible heat in the extra products of combustion.

b. The bed depth of the Carbon-Burnup Cell would either be fixed at the value of the primary cells (one fan for whole system) or, if a second fan were used, the bed depth in the Burnup-Cell would be increased until the gain in steam production due to increased combustion efficiency equaled the value of the additional fan power.

c. The firing rate (actually superficial velocity) must be roughly equivalent to that in the primary cells or a isolated Carbon-Burnup Cell would be required; e.g., if the Burnup-Cell's firing rate were 1/4 that in the primary cells the primary cell bed material could not be properly fluidized when it flowed into the Carbon-Burnup Cell.

d. The bed temperature would be maintained at the highest level consistent with preventing excessive agglomeration. Excessive agglomeration may occur at temperatures above about 2050°F where bed particle growth begins. If the bed particle growth were accommodated in the design (See Section 6.6.5) the limiting temperature would be the bed collapse temperature, possibly as high as 2400°F.

It will be shown later that the simple linear forms of the parameters are not appropriate for optimization.

What should be obvious from the preceding discussion is that a high bed temperature ( > 1900°F) is desirable. In the next section it will be shown how a high bed temperature is maintained.

### 6.4.5 Maintaining the Bed Temperature for High Combustion Efficiency: Effect of Coal Addition to the FBC

Figure 19 shows that if the Carbon-Burnup Cell is to operate efficiently at high throughputs with a shallow bed and moderate levels of excess stoichiometric air, the bed temperature must be maintained at a relatively high level. There are problems in maintaining the bed temperature at a high level in view of possible variations in the fly ash supply and its heating value which will occur in a commercial system.

One convenient method of maintaining a constant bed temperature is to add coal to the fuel fed to the burnup-cell at a rate sufficient to maintain the bed temperature at the desired steady state value. One important question to be answered was how coal addition would affect the combustion efficiency of the fly ash. A second question was how useful the mathematical model would be when part of the fuel was coal. Therefore, some experiments were run in which coal was fed with the fly ash. Four tests were run: C-314-4, C-319-3, 4 and 5. In test C-319-5, no fly ash was fed, only coal, to see if the mathematical model would work in the extreme.

Table 11 below presents the values of carbon combustion efficiency predicted by the mathematical model and the observed efficiency for these tests.

TABLE 11

PREDICTED AND OBSERVED CARBON COMBUSTION EFFICIENCY FOR CARBON-BURNUP CELL OPERATION WITH COAL IN FEED

| Test No. | Ratio Fly Ash Carbon/Coal Carbon | Carbon Combustion Efficiency, %* | | |
|---|---|---|---|---|
| | | Observed | Calculated | Residual |
| 314-4 | 3.03 | 89.5 | 88.5 | +1 |
| 319-2 | ∞ | 90.6 | 94.0 | -3.4 |
| 319-3 | 6.97 | 92.0 | 95.5 | -3.5 |
| 319-4 | 1.14 | 92.9 | 99.8 | -6.9 |
| 319-5 | 0 | 95.5 | 103.6 | -8.1 |

*Results not adjusted for hydrogen content of fuel.

It is seen that the actual carbon combustion efficiency is maintained at approximately 90% over a wide range of the ratio of fly ash carbon to coal carbon.

The mathematical model is less useful in predicting carbon combustion efficiency when coal is added than when fly ash alone is the fuel. However, the calculated and observed values may be adjusted by including the hydrogen in the fuel in the results. The adjustment consisted of treating two moles of hydrogen as if they were one mole of carbon, and using the adjusted carbon rate to predict an adjusted combustion efficiency. This is described in detail in Appendix I.

TABLE 12

ADJUSTED* VALUES OF COMBUSTION EFFICIENCY WHEN COAL IS FED TO A CARBON-BURNUP CELL

| Test No. | Ratio Fly Ash Carbon/Coal Carbon | Predicted and Observed Values of the Adjusted Combustion Efficiency for a Carbon-Burnup Cell | | |
|---|---|---|---|---|
| | | Observed | Calculated | Residual |
| 314-4 | 3.03 | 90.5 | 86.7 | +3.8 |
| 319-3 | 6.97 | 92.4 | 94.5 | -2.1 |
| 319-4 | 1.14 | 93.7 | 97.0 | -3.3 |
| 319-5 | 0 | 96.3 | 98.5 | -2.2 |

*See Appendix I for adjustment technique.

Table 12 shows that the statistically derived model is capable of predicting the combustion efficiency of a Carbon-Burnup Cell when coal is added to the fly ash, or even when coal is the sole fuel. Results for Test 314-4 should be disregarded since the 10" static bed used is too low for efficient fly ash combustion.

The requirements for accurate prediction are therefore that the carbon rates be adjusted and that combustion efficiency be defined on a heating value basis (See Appendix I). A further requirement is apparently that the bed depth be on the order of 18" or more. This latter requirement follows from the significance of bed depth to the combustion of fly ash and its lack of importance in coal combustion.

The use of coal as a method for maintaining Carbon-Burnup Cell bed temperature has been shown to be effective. High temperature in turn leads to high combustion efficiency in the Carbon-Burnup Cell. However, as the ratio (Coal fed to the Carbon-Burnup Cell):(Coal fed to Multicell Fluidized-Bed Boiler) increases, the overall combustion efficiency must drop. This follows from the fact that the coal fed to the burnup-cell has only one pass through the system. This may be understood by way of an example:

Assume that the data for tests C-319-2, 3, 4 and 5 represent the operation of a Carbon-Burnup Cell burning the fly ash collected from the flue gas of the primary cells of a Multicell Fluidized-Bed Boiler. Test 319-2 is assumed to represent full load operation; i.e., the Burnup Cell is operating at design capacity, and no coal need be fed to this cell. In tests 319-3, 4 and 5, the load on the primary cells is assumed to drop while the FBC (burnup-cell) operates at an essentially fixed air rate, 700 lbs/hr; thus coal must be fed to maintain the Burnup Cell bed temperature. If it is assumed that the primary cells operate at 90% combustion efficiency at all loads the overall combustion efficiency for this particular case would be as represented by the bottom curve in Figure 25. Note that the data points on Figure 25 apply to the case when the burnup cell is not capable of turndown. Other curves on Figure 25 are plotted for a burnup cell with turndown (i.e.. for the case when coal feed is not necessary when fly ash feed to the Burnup Cell is reduced).

The use of high-sulfur coal in the Carbon-Burnup Cell could lead to relatively high levels of $SO_2$ in the Carbon-Burnup Cell off-gas. This higher emission might not

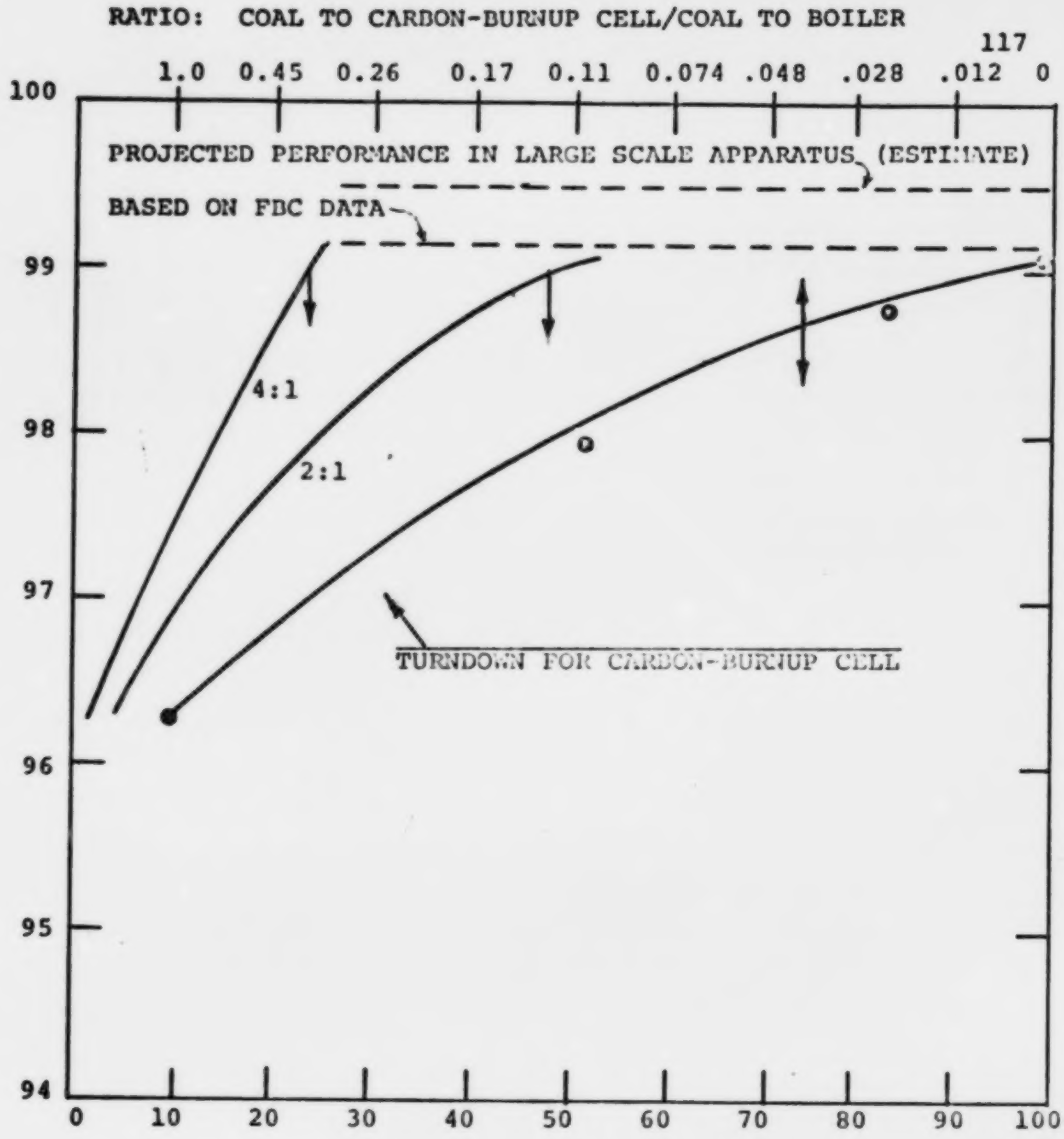


FIGURE 25 PREDICTED COMBUSTION EFFICIENCY OF MULTICELL FLUIDIZED-BED BOILER

be acceptable if when the CBC off-gas is mixed with the larger stream of boiler off-gas, the total $SO_2$ content of the combined gas stream exceeds the limit established by the local air pollution control authorities.

Two alternatives are available to insure that a fluidized-bed boiler in which fine limestone injection is used does not exceed the required $SO_2$ emission level.

1. Carbon-Burnup Cell bed temperature maintenance may be accomplished without the use of coal by the inclusion of a high temperature air heater in the boiler circuit. Under full-load conditions the air from the high-temperature heater would flow into the boiler circuit and the Carbon-Burnup Cell would receive ambient air. As Carbon-Burnup Cell bed temperature tended to drop, the high temperature air stream would be partially diverted to the Burnup-Cell in order to maintain the desired bed temperature. At the lowest fly ash feed levels, the Burnup-Cell would be receiving only high temperature air (approximately 1000°F).

2. Alternatively, Carbon-Burnup Cell off-gas could be wet scrubbed and the residual $SO_2$ removed by the lime also contained in the gas stream. The cold wet Carbon-Burnup Cell gas would then be mixed with the hot, dry and relatively sulfur-free boiler gas. It may be shown by a heat and mass balance that the mixed gas stream would be only 25°F cooler than the primary cell gas and gain very little additional humidity. The gas buoyancy would therefore be relatively unaffected.

A different set of alternatives are appropriate if sulfur emission control is achieved by using coarse limestone as the bed material and sulfur control agent, and/or if sintered ash by-products are to be produced in the Carbon-

Burnup Cell.

1. The fly ash carbon plus coal might be used as the reductant in a limestone regenerator. This alternative is presently under study and will be discussed in a future report.

2. If sintered ash by-products are to be produced, it is likely that collected fly ash would be stored so that the Carbon-Burnup Cell would be operated at a constant load regardless of the primary cell load. A wet scrubber would very probably be used for both $SO_2$ and particulate control.

### 6.4.6 The Effect of Fine Limestone Sorbent on Combustion Efficiency in a Carbon-Burnup Cell

Fine limestone may be injected into the primary cells of a fluidized-bed boiler to reduce emissions of sulfur dioxide. Limestone by-products would then appear in the fly ash feed to the Carbon-Burnup Cell. One original concern was that the presence of limestone by-products in the feed would adversely affect the combustion efficiency of the Carbon-Burnup Cell. At the beginning of the program, it was believed that the limestone would act as an inert diluent to the fuel and would have less adverse effect on combustion efficiency than the natural ash content of the fly ash -- if the excess oxygen in the gas were above about 3%.

A second concern was that the $CaSO_4$ in the fly ash feed might decompose at residual oxygens below 3 to 4% (See Section 6.6.4); the calcium sulfate produced in the boiler would tend to be decomposed at the high temperatures of a Carbon-Burnup Cell. In addition to releasing $SO_2$, the decomposition might be unfavorable to high combustion efficiency since the reaction

$$C + CaSO_4 + \frac{1}{2} O_2 \rightleftarrows CaO + SO_2 + CO_2 \qquad (12)$$

is endothermic. Note, however, that calcium sulfate represents an oxidant in the reaction and might increase carbon consumption. If substantial decomposition of calcium sulfate occurred in the Carbon-Burnup Cell, the flue gas from the Carbon-Burnup Cell would have to be treated to maintain the overall sulfur capture efficiency of the boiler.

Two tests, 320-1 and 320-2, were conducted in the FBC in which fly ash containing limestone by-products was the fuel. The fly ash for both tests had been collected during FBM test B-16 in which fine limestone had been injected. The collection of this material and its use in the FBC are potentially dangerous*, and so the number of tests was limited to two. Test 320-1 was run under strongly oxidizing conditions ($O_2$ = 4.3%) while test 320-2 was run under only mildly oxidizing conditions to encourage sulfate decomposition. This test is discussed later.

Table 13 below presents the key results of Test 320-1, the oxidizing test.

Table 13 shows that the $SO_2$ emission is only 350 ppm (about 6% of the sulfur input). Therefore, calcium sulfate does not decompose under the normal conditions of a Carbon-Burnup Cell; i.e., high temperature, high residual oxygen. It also appears that the mathematical model overpredicts the lowering of combustion efficiency by treating

---

*The fresh fly ash contained CaO, quicklime. It had to be removed hot from the hopper and placed in an air tight 55 gallon drum. When it was used in the FBC it is poured into an open hopper. During both these steps fines become airborne. Lab personnel wore respirators, gloves, etc. during these operations and were very uncomfortable.

## TABLE 13

## RESULTS OF FBC TEST 320-1 WITH FLY ASH CONTAINING LIMESTONE BY-PRODUCTS

| | Value |
|---|---|
| Air Rate, lb/hr | 620 |
| Bed Temperature, °F | 1980 |
| Bed Height (static), in. | 20 |
| Fuel Rate, lb/hr | 167 |
| Carbon Content, Wt. % | 27.9 |
| Carbon Rate, lb/hr | 46.5 |
| Inert Rate, lb/hr | 120.5 |
| Sulfur Content, Wt. % (Total organic & sulfate) | 2.42 |
| $O_2$ in Flue Gas (RO), vol. % | 4.3 |
| Calcium Content, Wt. % | 19.4 |
| Calculated Combustion Efficiency, % | 77 |
| Observed Combustion Efficiency, % | 87 |
| Residual, % | + 10 |
| $SO_2$ in Flue Gas, ppm | 350 |

the large limestone by-product load as ash. The combustion efficiency is approaching the desired 90% level and appears to be relatively unaffected by the high lime by-product load.

It is possible to adjust the inputs to the model by determining the inert load due to limestone by-products and deducting that amount from the total load of 120.5 lbs/hr indicated in Table 13 for the inert rate.

A detailed chemical analysis of the fly ash feed was not made to determine the amount of the various forms of calcium (CaO, $CaCO_3$ and $CaSO_4$). Instead the following is assumed: (a) 95% of the sulfur is in $CaSO_4$; (b) 20% of the remaining calcium (not in $CaSO_4$) is in $CaCO_3$; and (c) the remaining calcium (not in $CaSO_4$ and $CaCO_3$) is CaO.

These are reasonable assumptions. When this is done the inert rate, I, put into the model equation is 60.5 lbs/hr and the predicted combustion efficiency is 80.8%. The residual, +6.2%, is more deviant than any of the 38 tests used to generate the model.

These results make it appear (based on one test) that limestone by-products enhance combustion efficiency of carbon in a Carbon-Burnup Cell. What is more likely however considering the arduous conditions of this test, is that an error was made in recording either the fuel rate or the air rate. These two items of data are obtained at the FBC operating station. A second possibility is that the measured carbon content of the feed material (27.9%) includes some carbon as $CaCO_3$. If in fact 20% of the calcium not in $CaSO_4$ were in $CaCO_3$ the observed combustion efficiency of the elemental carbon would drop to 86.5% and the calculated value would rise to 82.3%. Unfortunately, a detailed analysis of the fly ash feed to determine the forms of carbon and cacium was not done.

For design the model may be used even with fine limestone present in the fly ash feed if the limestone is ignored in the calculations; i.e., treat it as containing two streams - a normal stream of boiler fly ash and another stream of fine sorbent ignored in employing the model.

The results of test 320-1 reinforce our finding that the inherent inert content of a normal fly ash is a measure of the availability of the carbon for further reaction, also see Curve D on Figure 23. The carbon of a fly ash high in carbon content is seen as being more reactive than a fly ash low in carbon content. Despite the possible flaws, the results of tests 320-1 lead to the conclusions that limestone by-products do not adversely affect the operation of a Carbon-Burnup Cell and of equal importance -- under sufficiently oxidizing conditions $CaSO_4$ does not decompose at the high temperature of a Carbon-Burnup Cell.

### 6.4.7 The Effect of Particle Size of Input Fly Ash

Relationships for burning time for small (100µm to 1000µm) particles of carbon take the form

$$t_b = Kd_o^2 \tag{13}$$

where

$t_b$ = burning time, seconds

$K$ = an empirical constant, 300 to 2000 sec/cm$^2$ (22)

$d_o$ = initial particle diameter, cm

Other, more complex forms, have been suggested in which terms for temperature and oxygen partial pressure are added.

For a Carbon-Burnup Cell, the time-for-burning is the time the carbon particle spends in the dense phase fluidized-bed plus the time the particle spends in the hot, oxygen-rich freeboard. This time will be about one second in a commercial Carbon-Burnup Cell. In the FBC the time-for-burning is somewhat less. On the basis of a one-second time-for-burning Equation 13 indicates, depending on the value for K, that particles over about 200µm to 600µm may not burn out in a Carbon-Burnup Cell. Although experimental data for particles less than 100 μm are not available, particles under 10μm should burn almost instantaneously in a Carbon-Burnup Cell.

In all FBC runs, the fuel feed was fly ash captured by a cyclone-type dust collector, since neither fly ash source was equipped with an electrostatic precipitator. The efficiency of a typical cyclone collector is relatively low for particles below 10 microns in diameter and approaches zero for particles under one micron. Therefore, the feed material to the FBC was depleted in the finest size fraction of the gas-borne dust produced by the furnace. This is shown in Figure 26 in which the distribution of the fly ash fed to the FBC Test in C-312 is shown as Curve A. An efficiency curve for a typical medium efficiency cyclone is also plotted on the figure as Curve B. in the collected sample 8 micron particles made up less than 1% of this sample although particles between 7 μm and 9 μm probably made up 2% of the gas-borne dust entering the fly ash source's collector.

The shaded area on the figure, marked 1, represents material that was not fed to the FBC, but which would have been had the Capitol Power Plant and the FBM been equipped with electrostatic precipitators.

The effect of this depletion on the performance model is to understate slightly the combustion efficiency which will be achieved in a commercial boiler where the entire primary cell dust stream is fed to the Carbon-Burnup Cell. The assertion that the combustion efficiency of the Carbon-Burnup Cell would increase had the fine fraction been included was not demonstrated experimentally. Instead the assertion was based on the assumption that the finer material would burn out more effectively, as implied by Equation 13.

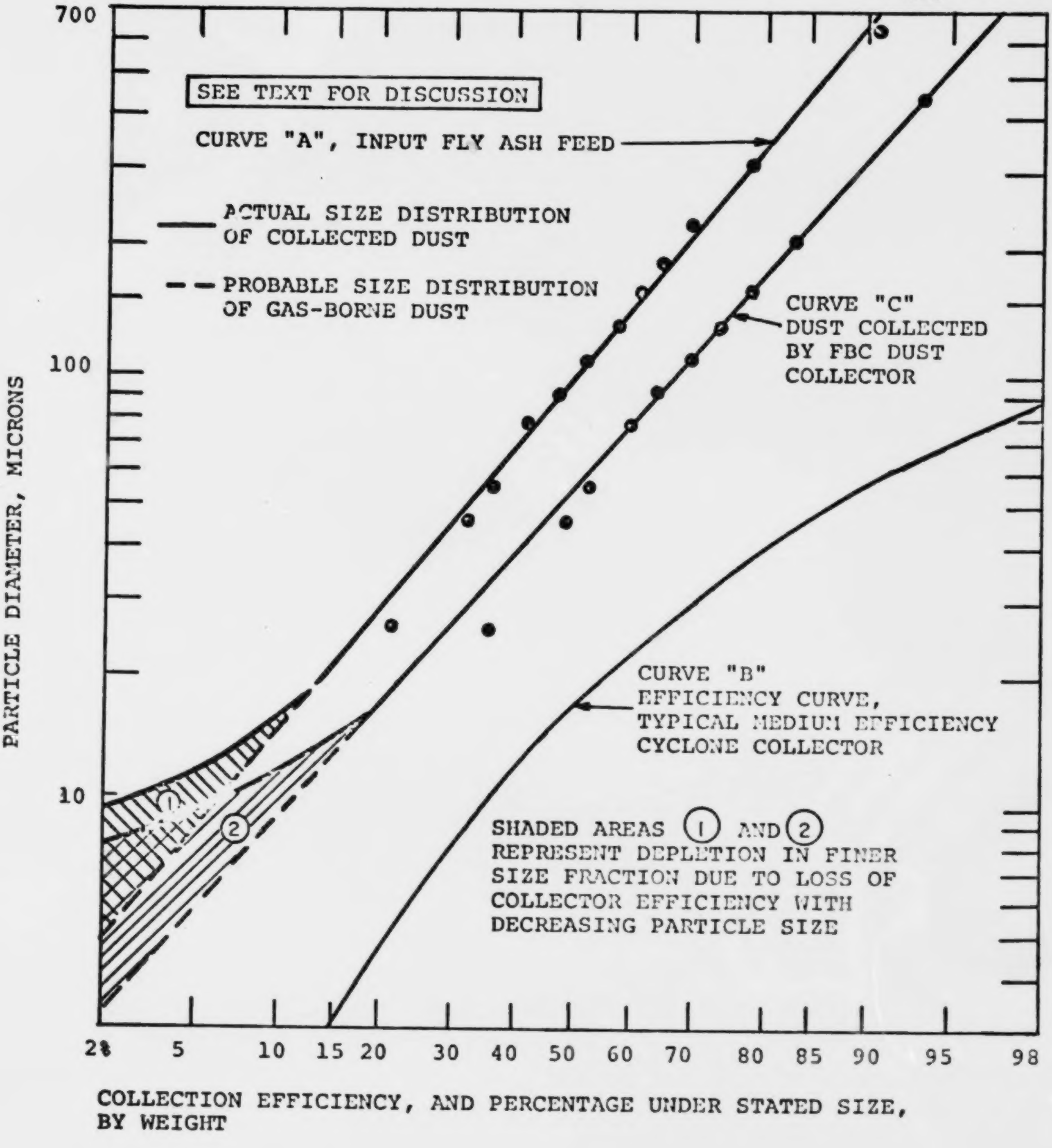


FIGURE 26 PARTICLE SIZE DISTRIBUTIONS FOR FBC TEST C-312

Data from an FBC test which might support this conclusion is shown below in Table 14, in which the feed and output for test C-314-2 were analyzed for carbon content versus particle size. Data similar to that used in preparing Figure 26 is not available for this test.

## TABLE 14

## CARBON CONTENT OF INPUT AND OUTPUT FLY ASH AS A FUNCTION OF PARTICLE SIZE

| (Data from Test C-314-2) | Carbon Content, % by Weight | |
|---|---|---|
| Sample Particle Size | Feed Fly Ash | Output Fly Ash |
| + 295 µm | 69.1 | 33.5 |
| -295 µm + 210 µm | 50.6 | 14.0 |
| -210 µm + 125 µm | 67.7 | 21.5 |
| -125 µm + 38 µm | 54.8 | 21.7 |
| - 38 µm | 31.4 | 9.7 |
| Average Carbon Content of Entire Sample | 56.0 | 18.6 |

Note that in the input material the lowest carbon content is found in the particles under 38 µm. After the fly ash was burned in the FBC the finest fraction had the lowest carbon content, but this is not the same material that entered the FBC as fines; i.e., the larger particles become small as they are burned.

The important information provided by Table 14 is that the carbon content of large output fly ash particles, +295 µm, is roughly twice that of the average for the entire sample. The definitive tests which might have answered the question: How does mean particle size affect combustion efficiency?, were not run since there was no real incentive to do so. It was already apparent from the data available that an overall combustion efficiency of 99% could be achieved by firing primary cell fly ash, without grinding or other preparation, into a properly designed Carbon-Burnup Cell. For test 314-2, the FBC operated with a 10" deep bed which is obviously too shallow, and achieved a combustion efficiency of 82%.

It would appear from Table 14 that particles above about 300 μm elutriating from the Burnup-Cell (these can be removed from a gas stream simply by a change in direction; e.g., by a heat exchanger baffle as in the FBM) might be recycled to the Carbon-Burnup Cell.

Size distribution data for the collected fly ash from FBC test 312-1 is also plotted on Figure 26 as Curve C. Although carbon content versus size data is not available from this test, it is reasonable to assume that it would be similar to that listed above for test 314-2; the overall combustion efficiencies were comparable. If these assumptions are valid then the fraction of particles larger than 295 μm, which is about 10% of the mass, contains about 20% of the unburned carbon for the effluent dust.

The material not collected by the FBC dust collector is also depleted in fines shown by the shape of Curve C on Figure 25. The size distribution of this material, indicated by the shaded region marked 2, was determined by sampling the gas leaving the cyclone with an isokinetic probe, and is plotted on Figure 27. The amount of material collected during test 312 on the isokinetic probe was sufficient only for the size analysis shown on Figure 27 and so carbon content data

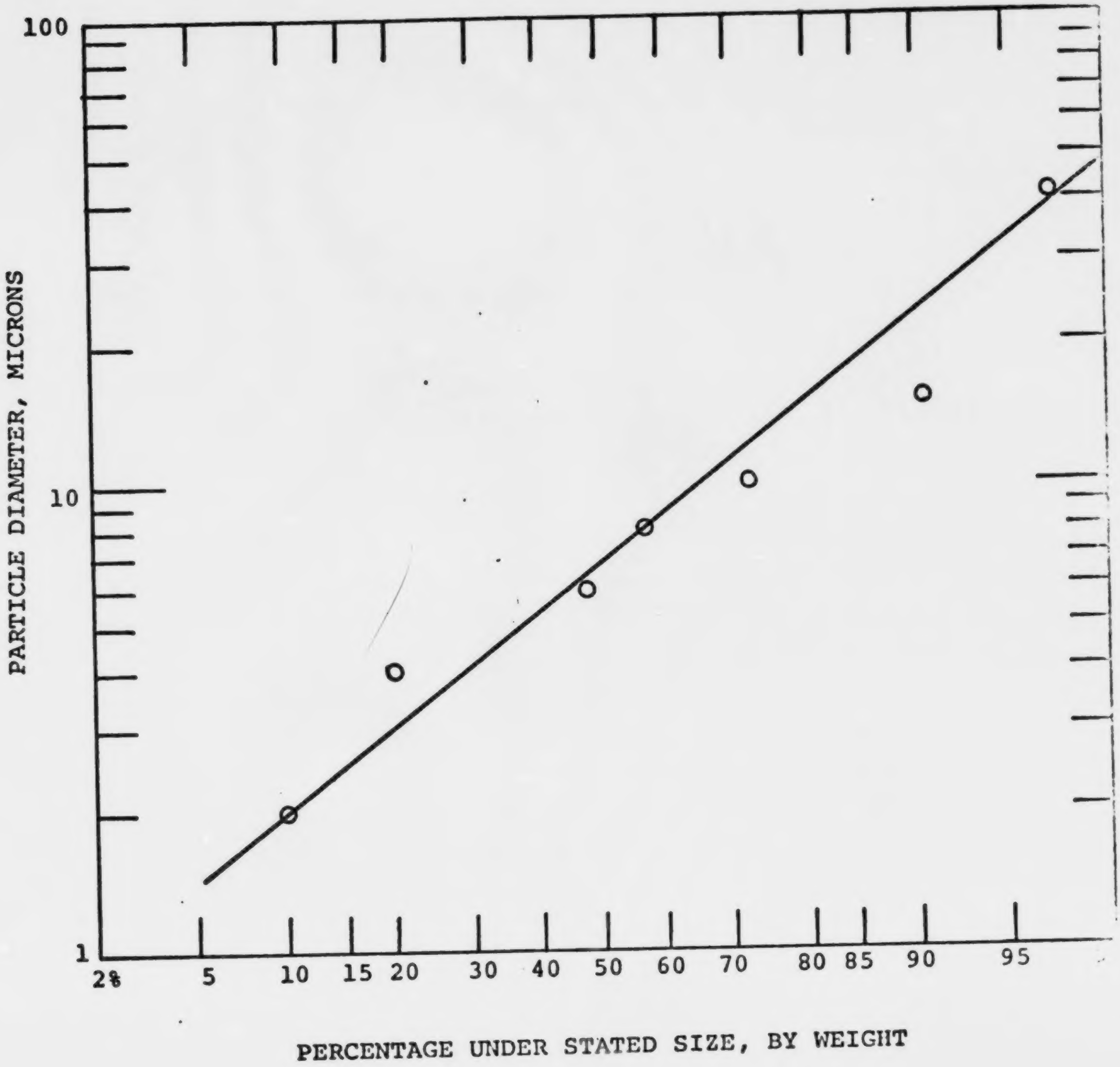


FIGURE 27 PARTICLE SIZE DISTRIBUTION OF FLY ASH DISCHARGED TO ATMOSPHERE - FBC TEST C-312

are unavailable for the particulate emission during the test.

On the basis of the material presented above the following conclusions are drawn:

(1) The performance model understates slightly, the combustion efficiency for a Carbon-Burnup Cell since the finest fraction of fly ash generated in the source furnace was not fed to the FBC.

(2) The coarsest fraction may not burn out in a single pass through a Carbon-Burnup Cell.

(3) If the primary cells are equipped with an electrostatic precipitator and if the coarsest fraction, (+300μm ) of dust from the Carbon-Burnup Cell is recycled, the combustion efficiency of the Carbon-Burnup Cell would increase by about 1 to 2% and of the Multicell Fluidized-Bed Boiler by 0.1 to 0.2%. An obvious alternative to recycling the coarse fraction would be to grind the fly ash feed to a fine powder so that all entering particles would be - 100μm. The beneficial effect of grinding the feed fly ash would have to be balanced against operating a pulverizer on a material' which is 50% (or more) inert. Since for Run C-312, the run for which the size data were analyzed, an 84% combustion efficiency was achieved with an untreated feed (mean particle size 100 microns), grinding of the entire fly ash stream does not appear warranted.

It is unfortunate that data on particle size distribution and carbon contents for each fraction were not obtained for each of the 38 tests used in modeling. Had such data been obtained it might have been possible to reduce the scatter in Figure 16; i.e., the model would be more accurate.

But, size distribution, just like overall carbon content, are not real control variables. They result from primary cell operating conditions; e.g., coal size consist. It might therefore be argued that the less precise model, that in which the effect of particle size is ignored, is the more useful design tool in commercial boilers.

### 6.4.8 Conclusions Regarding Statistically Derived Model for Carbon-Burnup Cell Performance

It has been shown that a model derived by the statistical analysis of the results of 38 fly ash combustion tests effectively predicts the carbon combustion efficiency of the FBC operating as a Carbon-Burnup Cell. The model is useful over the following range of values:

| | | |
|---|---|---|
| 1. | Bed temperature | 1750 to 2140°F |
| 2. | Bed depth (static) | 10 to 22 inches |
| 3. | Fuel rate (fly ash) | 60 to 300 lbs/hr |
| 4. | Air rate | 330 to 860 lbs/hr |
| 5. | Fuel carbon content | 28 to 65% by weight |
| 6. | Two bed particle size consists: | -16 +22 and -8 +16 mesh |

It was further shown that the model may be extended to include coal as part of the feed by applying the techniques indicated.

When large quantities of limestone by-products are included in the fuel stream, the model is less useful even after the feed parameters are adjusted for the diluent. However, even here the difference between the adjusted calculated value and the observed value is within the range required for a design correlation. It was not found possible to produce a correlation when heat removal was substituted for bed temperature as an independent variable in the model.

However, it will be shown later that there is a reasonably well-defined relationship between heat removal and combustion efficiency.

The accuracy of the model would probably have been improved if particle size data had been obtained and input to the model.

The trends for the parameters are clear and agree with intuition. Combustion is more complete at higher temperatures, with deeper beds, with lower throughputs and with higher ratios of air to carbon.

It is possible to use the model equation to design a Carbon-Burnup Cell if plan area is the appropriate scaling factor (this is discussed in the next section). It does not appear possible to use the equation to optimize the design. This flaw follows from the lack of curvature terms* in the equation. The model will therefore predict a combustion efficiency greater than 100% (See Appendix F for a discussion of the statistical consultant's attempts to correct this flaw).

The combustion efficiency of carbon, in fly ash, in a Carbon-Burnup Cell does appear to increase linearly with each parameter from low values to the point where CE $\simeq$ 90%. Beyond where CE $\simeq$ 90%, the effect of increasing bed temperature (or bed depth, etc.) should be diminished by a square term with a negative coefficient, thus

$$CE = b_0 + b_1 \ (T - a_1 T^2), \text{ etc.}$$

---

*Curvature terms comprise square terms plus second order interaction terms other than C x I.

Such a form could not be derived from the small amount of data in the high CE range. The FBC was simply not big enough, adiabatic enough and tall enough to explore the regime near CE → 100% at the throughputs which are of commercial interest.

Without curvature terms, there is no basis for trade-offs. For example, if CE = 90% with a bed depth, H, of 22 inches, the model indicates that CE = 100% will be achieved if the depth were increased to 48", all other parameters constant. It may be shown that for reasonable values of A/C, the air-to-carbon ratio, more energy is made available by burning the extra 10% of the carbon than is used in driving the bigger fan required by the deeper bed. But this is unreasonable; the relationship between combustion efficiency and bed depth cannot remain linear as CE → 100%.

Despite the flaws listed in the preceding paragraphs, the Carbon-Burnup Cell performance model is judged to be a useful design tool.

## 6.5 Applying the Performance Model in Design

### 6.5.1 General

The mathematical model which predicts carbon combustion efficiency was derived from the statistical analysis of 38 tests conducted in the FBC. While the range of parameters studied was remarkably wide (only bed temperature did not vary by a factor of two or more), only one apparatus was used. There is no absolute certainty that some change in the design of the apparatus would not have affected the results markedly.

If the apparatus had been conical rather than straight-sided, the coefficient for bed depth might have been totally different. While bed particle size was not found to have a profound effect on the results, only sintered coal ash was evaluated, and so a change to, say, alumina or sulfated lime might have affected the results.

A possible problem in any statistical correlation, such as this one is that its extension to conditions and/or apparatus not studied may prove unsuccessful.

The alternative would be a physical model which describes, analytically, the entire process of fluidization and combustion. Given the current status of fluidization theory and heterogeneous combustion theory, a useful physical model is not available.

Despite the above, the statistical model is felt to be valid and extendible because of the range of parameters over which it was derived and because the apparatus was large enough (and air rates high enough) to minimize wall effects.

The first step in applying the model in design is to normalize the model, derived in the FBC, so that it may be used for the much larger systems.

### 6.5.2 Normalizing the Model for Use in Other Apparatus

The Carbon-Burnup Cell performance model, Equation 6, was derived from tests in the FBC. The data used in the derivation were presented in the simplest possible format; i.e., as they were measured in units such as pounds per hour, degrees fahrenheit and inches of static bed depth. The performance model as given by Equation 6 requires information to be input in this same format. It should be obvious that each parameter has the implicit dimension -- per "FBC". So the air rate should be expressed as pounds of air/hour - "FBC"; bed depth as inches of depth per "FBC" and possibly even temperature as $^{o}$F per "FBC". In the regression coefficients presented in Table 9 there is also the implicit dimension -- "FBC". Thus for CE; $b_0$ = -13.78 "FBC", $b_1$ = +0.05193 "FBC"/$^{o}$F, $b_2$ = +0.0462 "FBC" -hr/lb, etc.

To normalize the Carbon-Burnup Cell performance model it is necessary to express the dimension, "FBC", in units of mass, length and time. This is the common problem of determining scaling factors and a different scaling factor may be required for each parameter.

This is a serious problem, and although a scaling factor was selected, as described below, the basis was arbitrary. For this reason the model equation and coefficients were left in their original format until this point in the study.

The basic scaling factor selected is the simplest, plan area, and affects each regression coefficient as follows:

For $b_0$ (a constant) an "FBC" = 1 (dimensionless)
$b_1$ (for bed temperature) an "FBC" = 1 (dimensionless)
$b_2$ (for air rate) an "FBC" = 0.86 ft$^2$
$b_3$ (for bed depth) an "FBC" = 1 (dimensionless)
$b_4$ (for carbon rate) an "FBC" = 0.86 ft$^2$
$b_5$ (for inert rate) an "FBC" = 0.86 ft$^2$
$b_6$ (for carbon rate inert rate cross product) an "FBC" = 0.76 ft$^4$

It is believed that these scaling factors will give valid results if it is understood that the carbon rate given in pounds of carbon per hour per square foot of grate area must be evenly distributed over this area. It is believed, but not proven, that the model will be valid for a point injector supplying a grate area of about $\pi$ $ft^2$.

The normalized coefficients will now be derived for the performance model. It will be useful to do this by way of a sample calculation for a test in the FBM. The FBM test used is Test B-14, for which the air rate to the CBC was 1227 lb/hr $ft^2$. The sample calculation is given on page D-2. No adjustment is required for bed depth or temperature.

The normalized form of the performance model, presented differently in the summary, is

$$CE = b_0 + b_1 (T) + b_2' (A) + b_3 (H) + b_4' (C') + b_5' (I') + b_6' (C' \times I') \qquad (1A)$$

where

$b_0 = -13.78$ (Dimensionless)
$b_1 = +0.05193$ ($^oF^{-1}$)
$b_2' = +0.03973$ ($ft^2 hr/lb$)
$b_3 = +0.3831$ ($in^{-1}$)
$b_4' = -0.7514$ ($ft^2$ $hr/lb$)
$b_5' = -0.1638$ ($ft^2$ $hr/lb$)
$b_6' = +0.0020$ ($ft^4$ $hr^2/lb^2$)

Applying the normalized data to the normalized performance model (Equation 1A), the predicted combustion efficiency for FBM test B-14 is 95.8%. The observed value was 95%. The residue is -0.8%.

When FBM test B-15 was analyzed, in which the air rate to the CBC was 1880 lb/hr, the predicted value of combustion efficiency was 117% against an observed value of 93%. (The model contains no provision to prevent values over 100%, See Appendix F). It appears, therefore, that the model, based on tests with a maximum air rate of 1000 lb/hr per ft$^2$ (860 lb/hr into the 0.86 ft$^2$ FBC), may be extended to an air rate on the order of 1250 lb/hr ft$^2$ (25% extrapolation), but it is useless at air rates of 1700 lb/hr ft$^2$ (70% extrapolation); i.e., the model needs a parabolic term for air rate so that it is not linear at high air rates.

A comparison of observed and calculated values for CBC tests with lower air rates is given on Figure 37 in Section 7. Some of the residuals, especially for the earlier tests, are high; e.g., for test B-5 the residual = +16, an absurd value. For test B-12 on the other hand where the air rate was 750 lb/hr ft$^2$ the residual = 0.

The following conclusions are drawn regarding the application of the Carbon-Burnup Cell performance model to larger apparatus and conditions outside the range used in the derivation:

1. The model is useful for devices 50% larger than the FBC.

2. The model is probably useful for rates* 25% to 50% higher than used in the FBC.

### 6.5.3 The Bed-Temperature Anomaly

In the preceding section a normalized form of the performance model was discussed. In the next section the use of this model in design will be illustrated. But first the problem of bed temperature prediction must be discussed.

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*Air, fuel, heat removal, etc.

It was noted in Section 6.2.5 that it is not possible to predict the bed temperature of a Carbon-Burnup Cell on the basis of the three parameters (1) fuel rate, (2) air rate and (3) a heat loss term expressed as Btu/hr°F which includes heat transfer coefficients and areas; or at least the attempt made here to do so was unsuccessful. Even when a fourth parameter, combustion efficiency, is added it is still not possible to predict bed temperature. It was further noted that this problem probably arises from three factors:

(1) the combustion process in the Carbon-Burnup Cell is bi-stable

(2) not all of the energy release, implied by a knowledge of the combustion efficiency and fuel rate, is released within the dense phase fluidized-bed and

(3) the heat loss to surfaces above the bed may be higher than anticipated.

These three problems are discussed below.

### 6.5.3.1 Bi-Stable Combustion

During testing (See Table 15, A) it was found possible to perform two experiments, the first in which the bed temperature was 2100°F and a second at a bed temperature of 1800°F. When the two experiments were complete and the data reduced it was found that the air rate, carbon rate, inert rate and bed depth were almost identical in both experiments. Since no cooling coil was used and the loss of heat to the enclosure was almost identical in both experiments, the bed temperature would have been expected

## TABLE 15

## ILLUSTRATION OF BI-STABLE COMBUSTION IN FBC

| Test No. | Air Rate, lbs/hr | Carbon Rate, lbs/hr | Bed Temperature, °F | Combustion Efficiency, % |
|---|---|---|---|---|
| A. Variation in Bed Temperature at Constant Feed Rate(1) | | | | |
| C-306-1 | 600 | 41.5 | 2100 | 85.5 |
| C-306-4 | 600 | 43.0 | 1800 | 72.7 |
| C-307-1 | 600 | 40.6 | 1800 | 76.0 |
| C-307-3 | 600 | 44.0 | 2140 | 91.0 |
| B. Variation in Carbon Feed Rate at Constant Bed Temperature | | | | |
| C-308-1 | 600 | 37.8 | 1900(2) | 81.5 |
| C-308-2 | 600 | 42.4 | 1900(3) | 81.3 |
| C-308-3 | 600 | 43.0 | 1900(4) | 80.6 |

[1] In the two test pairs there was a 3.5 to 7.5% variation in carbon feed rate but note that in 306 the higher temperature was at the lower carbon rate while in 307 the higher temperature was found at the higher carbon rate.

[2] Bed temperature reached by slow rise to steady state.

[3] Bed temperature reached after 15 minutes at 2100°F.

[4] Bed temperature reached after 15 minutes at 1500°F.

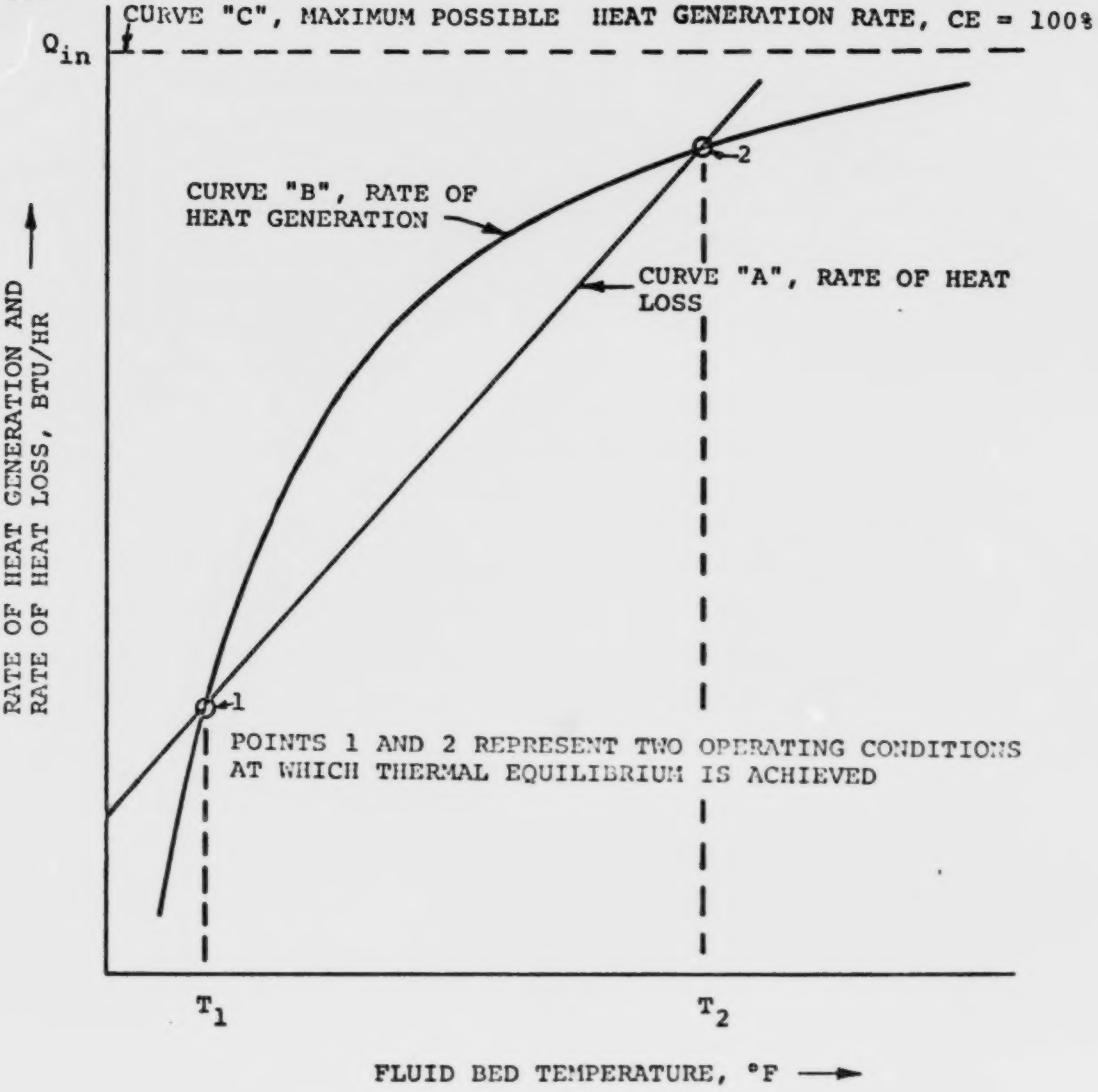
to be the same. That the system may perform differently under apparently identical conditions appears at first to be anomalous.

One possible explanation for the apparent anomaly is that the data were incorrect, but in a second set of experiments the anomalous results were again observed. It was also suggested that the bed might be "conditioned" by its previous temperature history. Tests 308-1, 2 and 3 (see Table 15 B) with different preconditions showed that this was probably not a valid hypothesis. A more suitable explanation is that a stable "flame" could be achieved at more than one bed temperature, all other system parameters kept constant. This may be understood by way of a hypothetical illustration. Curve A in Figure 28 plots the energy lost by the bed as a function of the bed temperature. Curve A is a straight line based on the assumptions that: (1) the major fraction of energy lost by the bed is the sensible heat in gas and dust and that the enthalpy rise of these substances is proportional to the bed temperature and (2) the heat lost by the bed to the walls of the combustor is proportional to the bed temperature.

Curve B of Figure 27 is a plot of the energy released within the bed, also as a function of bed temperature. The shape of the curve is taken as the top portion of the classic S-shaped curve for a combustion process. The shape can be explained as follows:

(1) At some temperature below $T_1$, ignition occurs, the rate of combustion of carbon then increases exponentially with increasing temperature in the region where the process is reaction-rate limited. At higher temperatures, above that at which the process becomes diffusion limited, the rate curve increases more slowly with temperature since the effect of temperature on the diffusion rate is not so great. (The temperature at which the combustion process changes from being reaction rate limited to being mass transfer limited depends on a number of factors), or,

(2) Similar to (1) above except that the curve flattens at relatively high temperature simply because the



RATE OF HEAT GENERATION
= COMBUSTION RATE (LBS OF CARBON PER HOUR/FT$^2$ OF CARBON SURFACE IN BED) X CARBON SURFACE (FT$^2$ OF CARBON SURFACE IN BED) X HEAT OF REACTION (BTU/LB OF CARBON BURNED)

RATE OF HEAT LOSS
= ENTHALPY RISE OF PRODUCTS OF COMBUSTION + ENTHALPY RISE OF INERT PASSING THROUGH BED + HEAT LOSS THROUGH WALLS AND FROM BED SURFACE

FIGURE 28 ILLUSTRATION OF BI-STABLE COMBUSTION PROCESS

rate of heat generation cannot exceed the input rate $Q_{in}$.

The rate of heat generation in a fluidized bed may be seen to depend on the combustion rate expressed as pounds of carbon consumed per hour per square foot of carbon surface, the surface area of carbon existing in the bed at any time, and the heat of the reaction. The surface area of carbon existing in the system depends, on the one hand, on the rate at which carbon is being fed, and on the other, on the rate at which it is being lost by combustion and elutriation, and also on the size distribution of the carbon-containing particles within the bed. The temperature of the bed is that temperature at which the rate of heat generation is equal to the rate of heat loss. Two such temperatures are seen to exist on Figure 28. Point 1, represents the the thermal balance at a "low" temperature and point 2 at a "high" temperature.

In a pulverized coal furnace point 1 of Figure 28 would represent an unstable condition. Note that if a perturbation slightly increased the rate of heat generation the temperature would tend to increase. This in turn would increase the heat release rate and the system would move to point 2. If, on the other hand the combustion rate dropped slightly the temperature would also have to drop decreasing the heat generation rate such that the fire must go out.

Point 1 may be stable in a fluidized-bed combustor because:

(1) The thermal inertia of the bed allows the system to override minor perturbations, and

(2) Figure 28 is drawn only for the Bed, not the entire system which includes combustion in the freeboard.

When the combustion rate in the bed drops as it might at Point 1, the combustion rate may increase in the space above the bed. Energy released in the freeboard is returned to the bed most probably in the form of the sensible heat content of the refluxing bed particles. The energy partition between bed and freeboard is discussed below.

#### 6.5.3.2 Effect of Energy Release Patterns on Temperature of Fluidized Bed

It was shown in the preceding section that the fluidized-bed of a Carbon-Burnup Cell could achieve a thermal balance at two distinct bed temperatures. At the higher bed temperature, a larger fraction of the fuel which enters the bed burns within the bed than would burn there at the lower temperature.

It follows therefore that at the lower bed temperature more fuel appears in the freeboard as unburned carbon or carbon monoxide. Assuming that the air rate remains constant, as it did in the construction of Figure 28, oxygen must also exist in the freeboard and so some fuel will burn in the freeboard.

In a test run in the FBC it would not be possible to determine what fraction of the carbon burned within the bed and what fraction of the carbon was consumed in the freeboard. The combustion efficiency, derived from a carbon balance, was for the combustor as a whole.

The partition of energy release, between bed and freeboard may have a profound effect on the temperature of the bed as will be shown by the simple illustration which follows. Figure 29 has been prepared for a hypothetical Carbon-Burnup Cell and experiment to which the following

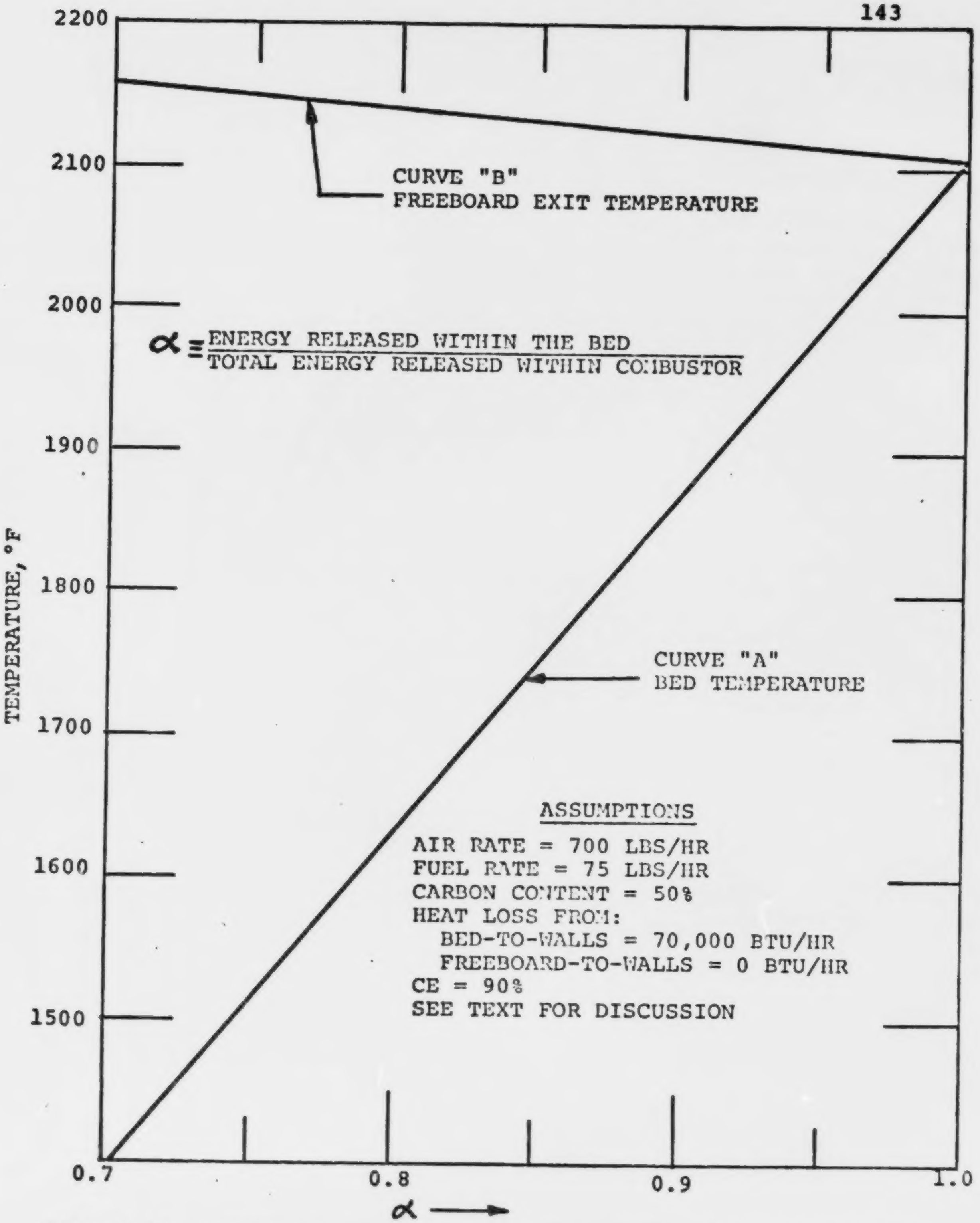


FIGURE 29 EFFECT OF ENERGY RELEASE PARTITION, α, ON TEMPERATURES HYPOTHETICAL FLUIDIZED-BED COMBUSTOR

data will apply:

| | |
|---|---|
| Air Rate | 700 lbs/hr |
| Fuel Rate | 75 lbs/hr |
| Carbon Content | 0.5 lb carbon/lb fuel |
| Heating value of carbon | 14,100 Btu/lb |
| Combustion efficiency | 90% |

The combustor has the following characteristics

(a) Heat loss from bed to walls, = 70,000 Btu/hr (This was typical of the FBC where heat loss to walls was almost independent of bed temperature)

(b) Heat loss from freeboard to enclosure or bed = 0 Btu/hr.
(This assumption is invalid but is useful for the sake of the illustration)

(c) Size of combustion chamber = Variable (The chamber changes as required to permit CE = 90% regardless of bed temperature).

Curve A in Figure 29 represents the bed temperature, Curve B the temperature at the exit from the freeboard for values of a parameter, $\alpha$ , where

$$\alpha = \frac{\text{Energy released within the bed}}{\text{Total energy released within combustor}} \qquad (14)$$

The system described in Figure 29 may be viewed as the extreme case in which the temperature of the bed is totally independent of the temperature in the freeboard. It is possible for such a system to exist, for example if fine coal were fed to a single point at the base of a low velocity ($\simeq$ 2 fps superficial velocity) fluidized-bed combustor contained in a large refractory enclosure. In such a case volatiles and coal fines would enter the freeboard and burn there. Exit gas temperatures might exceed 2500°F.

Data obtained in the FBC tests cannot be used to calculate values for $\alpha$ . The marked temperature differences between bed and freeboard implied by Figure 29 were not recorded. If it had been possible to determine the average gas composition at the "top" of the expanded fluidized bed it might have been possible to determine values for $\alpha$ as a function of the control variables. This was beyond the scope of the test program described in this report.

The discussion here was not intended to obtain values for $\alpha$ but to show a few of the implications of such a parameter. It can also be seen that for the FBC where the high air rates cause large quantities of bed materials to be thrown into the freeboard that a mechanism exists for returning energy released in the freeboard to the bed; i.e., in the sensible heat of the refluxing particles.

### 6.5.3.3 Energy Loss to Cool Surfaces Above the Dense Phase Fluidized Bed

In the preceding section it was noted that the bed of a Carbon-Burnup Cell can receive energy as sensible heat of particles which return to the bed at a higher temperature than the temperature at which they were initially expelled from the turbulent bed. This might occur if the freeboard were well insulated and if a fraction of the carbon feed burned in the freeboard instead of within the dense phase bed. If the freeboard is not so well insulated, the bed may lose energy to the cool surfaces above the dense phase bed.

Part of the inability to predict bed temperature for the FBC follows from the inability to measure or compute the energy lost, by the bed, to cooler surfaces above the bed.

In the FBC the cool surfaces above the bed included the upper part of the bayonet cooling tubes, the hood and the exhaust duct. In the FBC, it was not even possible to measure the energy lost from the dense phase bed to the walls in contact with the bed; the walls received energy from both the bed and the freeboard and only the total energy input to the walls was measured.

The transfer of heat from a fluidized-bed to heat transfer surfaces above the top of the bed does not appear to have been treated in the fluidization literature; one reason being the difficulty of getting useful data with small scale apparatus. But, despite the lack of quantitative data on the loss of energy by the bed to cool surfaces above the bed it is necessary to demonstrate here that such a loss does sometimes exist, and to indicate how its magnitude has been estimated. For purposes of illustration, assume that a fluidized-bed combustor is contained within a tall cyclindrical column. The wall of the column will be maintained at a temperature lower than the bed temperature. Under these conditions energy will be removed from the bed according to

$$Q_r = A_b \sigma (\epsilon_b T_b^4 - \alpha_w T_w^4) \qquad (15)$$

where

$Q_r$ = Radiative heat loss rate from fluidized bed to walls, Btu/hr

$A_b$ = Surface area of top of fluidized bed

$\sigma$ = Stefan-Boltzman's constant = $0.1713 \times 10^{-3}$ Btu/hr $ft^2 {}^oR^4$

$\epsilon_b$ = Emittance of fluidized bed $\simeq 1.0$

$\alpha_w$ = Absorptance of enclosure walls; $0.8 \leq \alpha_w \leq 0.95$ for typical surface walls

$T_b$ = Temperature of bed near surface $\simeq$ average bed temperature, ${}^oR$

T = Temperature of enclosure wall, $^{o}F$

When $(T_b/T_w)^4 << 1$ and assuming $\varepsilon_b = \alpha_w = 1$

Equation 15 becomes

$$Q_b = A_b \sigma T_b^4 \qquad (15A)$$

For

$T_b = 2000^{o}F\ (2460^{o}R)$

$Q_b = 62{,}600 \frac{Btu}{hr\ ft^2} \times (A_b,\ ft^2)$

The energy loss due to radiation to surfaces above the top of the bed could exceed the loss to submerged surfaces in a number of practical situations. To use Equation (15A) in predicting the bed temperature, it is necessary to estimate $A_b$, the area of the top of the fluidized bed.

If the fluidized bed were relatively quiescent, the surface area of the top of the bed, for the vertical cylindrical chamber would be

$$A_b = A_p = \pi R^2 \qquad (16)$$

where

$A_p$ = Area of flat plate

R = Radius of vessel

Where the bed is fired at high rates; e.g., approaching $10^6$ Btu/ft$^2$hr, the surface would be far from quiescent. It may be shown that the greatest surface area, for radiation, of the most irregular surface is given by

$$A_e = (\frac{\pi}{2}) \times A_p;\quad A_e/A_p = \pi/2 = 1.57 \qquad (17)$$

where

$A_e$ = Extended surface area (for radiation)

If the diameter of the fluidized bed were small enough, the bed would slug at high throughputs and the extended surface area of the top of the bed would approach the area of the absorbing enclosure; i.e.,

$$A_e \quad \text{(in the limit)} \rightarrow /2_{\pi}\ RH_{slug} \qquad (18)$$

where

$H_{slug}$ = Effective height to which slugs of bed material are thrown from the bed and are still close to the bed temperature.

Even where the bed does not slug, there may be sufficient material thrown into the freeboard to require the extended surface to be computed thus:

$$A_e = 2^{\pi}\ R\ H_{eff} = P\ H_{eff} \qquad (19)$$

where

$H_{eff}$ = Effective height to which particles are thrown from the bed in sufficient concentration to approximate a black body radiating at the bed temperature

$P$ = Perimeter of vessel, ft.

In preparing Figure 16, Equation 17 was used to compute the radiative loss from the fluidized bed. In previous (unpublished) work it has been found that the measured bed temperature of the FBM could be reproduced by an empirical model in which Equation 19 was used when $H_{eff} =$ 3.5 ft and the perimeter follows from the rectangular dimensions of the enclosure.

It follows from the definition of $H_{eff}$ that for very small R, $A_e/A_p \rightarrow 1$, and for very large R, $A_e/A_p < \pi/2$. At intermediate values of R, $A_e/A_p$ reaches a maximum. For the FBC, $A_e/A_p$ may have exceeded 10; no attempt to measure a real value could be made in the FBC.

The anomalous behavior of bed temperature stems in part from the inability to partition the energy absorbed by the hood of the FBC into two components; one primarily convective having small effect on bed temperature and a second primarily radiative having large effect on bed temperature.

### 6.5.4 An Example: Using the Carbon-Burnup Cell Performance Model in Design

Assume a Multicell Fluidized-Bed Boiler is to be designed to provide 250,000 lb/hr of steam at 600 psig, 750°F for the sample design calculation which follows. The performance specification contains an efficiency guarantee which will require 99% combustion efficiency at full load.

The coal to be used, at the rate of 27,500 lb/hr, has the analysis shown below in Table 16. It will be assumed that just enough of the coal ash will be retained in the bed to compensate for attrition; i.e., all the ash eventually appears in the fly ash. Fine limestone will be injected into the primary boiler cells to reduce emissions of $SO_2$ by 80%; based on past work this level of sulfur removal requires a Ca/S ratio of 2.5 and this value is assumed here. The primary cells will operate so that 3% residual oxygen appears in the stack gas. At this residual oxygen level, about 10% of the input fuel energy value normally appears as carbon in the fly ash. The value 10% will be used here. The dust collected in the primary cyclone and precipitator is to be fired in the Carbon-Burnup Cell of the Multicell Fluidized-Bed Boiler.

## TABLE 16

## COAL ANALYSIS FOR FUEL TO FLUIDIZED-BED BOILER, AN EXAMPLE OF CARBON-BURNUP CELL DESIGN

### ULTIMATE ANALYSIS

| Constitutent | %, By Weight |
|---|---|
| Carbon | 69.4 |
| Hydrogen | 4.7 |
| Nitrogen | 1.3 |
| Oxygen | 7.3 |
| Sulfur | 3.2 |
| Water | 5.0 |
| Ash | 9.0 |
| Total | 100.0 |

Btu/lb as fired = 12,550

To insure that no calcium sulfate is decomposed, the Carbon-Burnup Cell will operate at a residual oxygen level of 6%. Other cell design parameters are as follows:

Static bed height, H = 24", Bed temperature, T = 2050°F, Cell length, L = 10' (based on design arrangement).

Equations 1 and 2, restated below, and $C + O_2 = CO_2$ will be used in these computations in addition to a straightforward heat balance.

$$\begin{aligned} \text{CE, \%} = {} & -13.78 = 0.05193(T) + .03973\ (A') \\ & + 0.3831\ (H) - 0.7514\ (C') - 0.1638\ (I') \\ & + 0.0020\ (C'\ I') \end{aligned} \tag{1A}$$

$$\begin{aligned} \text{RO, \%} = {} & +22.91 - 0.007353\ (T) + 0.01118\ (A') \\ & - 0.1390\ (H) - 0.1521\ (C') - 0.0151\ (I') \\ & +.0002\ 5\ (C'\ I') \end{aligned} \tag{2A}$$

$C + O_2 = CO_2$ is rewritten below as a mass balance assuming the oxygen is supplied as air and using the nomenclature of Equation 1 and 2.

$$C(s) + \frac{CE/100}{(1.0-4.76x(RO/100)} \times O_2(g) + 3.76\ N_2(g) =$$

$$(CE/100) \times CO_2(g) + (1.0 - CE/100) \times C(s) + \qquad (20)$$

$$\frac{CE/100}{(1.0 - 4.76 \times (RO/100)} \times 4.76 \times (RO/100) \times )_2(g)$$

$$+ 3.76\ N_2(g)$$

where

$C(s)$ = moles of carbon fed as a solid

$O_2(g) + 3.76\ N_2(g)$ = molar composition of air

From the information given above, the following is known or may be computed:

CE = (99-90)/0.1 = 90, RO = 6, T = 2050$^{\circ}$F
H = 24", L = 10', A' = A/(L x W)
I' = I/(L x W), I = 27,500 x (.09) = 2475 lbs/hr
C' = C/(L x W), C = 27,500 (12,500)(0.1)/14,100
= 2,447 lbs/hr

Equations 1 and 2 may be solved simultaneously for two unknowns, A' and C'. Then the cell width, W, may be readily computed from W = C/(C' x L). Using this procedure C' = 69 lbs/hr ft$^2$, A' = 1,053 lbs/hr ft$^2$.

Alternatively Equation 1 and 20 may be solved simultaneously. If instead Equation 20 is used as a check for the air rate substitute C(s) = 69/12 = 5.75 moles/hr and the A' (Equation 20) = 1,012 lbs/hr ft$^2$; a reasonable agreement.

Next compute the cell width:

$$W = \frac{2447}{69\text{x}10} = 3.54 \text{ ft}$$

The final task is to perform a heat balance for the Burnup Cell and determine energy surplus (to be removed by tubes inserted in the bed) or deficit (to be made up by added coal).

The following information will be assumed as given:

(1) The walls of the cell will be made up of boiler tubes filled with boiling water at 645 psig (485°F).

(2) The expanded bed depth is 1.2 times the static bed depth.

(3) The heat flux into wetted surface is given by:

$q_w$ = U, Btu/ft$^2$hr$^\circ$F x ( $\Delta$T , $^\circ$F) = 50(2050-485)

= 78,200 Btu/hr ft$^2$

(4) The heat loss due to radiation to the enclosure is given by $Q_r$ = 66,500 Btu/hr ft$^2$ x $A_b$

(5) The radiating top surface of the fluidized-bed,

$A_b$ = $\pi$/2 (L x W) = 55.6 ft$^2$

in that $Q_r$ = 3,700,000 Btu/hr

(6) The combustion air is supplied at 600°F.

(7) Limestone by-products are fed to the Burnup Cell with the normal fly ash = 5,815 lbs/hr (assuming limestone is 97% $CaCO_3$, that all is calcined in primary cells, Ca/S = 2.5 and 80% of coal sulfur is removed as $CaSO_4$).

(8) Gas leaving the bed at 2050°F has an enthalpy value of 515 Btu/lb.

(9) Dust leaving the bed has an average enthalpy rise of 365 Btu/lb based on an average specific heat of 0.22 Btu/lb °F and an inlet temperature of 400°F.

(10) The $\alpha$ factor, defined by Equation 15, = 0.9.

| | | 1,000's of Btu/hr |
|---|---|---|
| A. Input | | |
| 1. From combustion within bed ( $\alpha$ = 0.9) | = | 28,000 |
| 2. From preheated combustion air | = | 4,700 |
| Total energy input | | 32,700 |
| B. Output | | |
| 1. In gaseous products of combustion | = | 20,800 |
| 2. In dust leaving bed | = | 3,200 |
| 3. To boiler tubes contacted by bed | = | 5,100 |
| 4. To boiler tubes viewed by bed, Q | = | 3,700 |
| Total energy output | | 32,800 |
| C. Net surplus (deficit) | = | (100) |

Based on the assumptions there is an energy deficit of 100,000 Btu/hr if the Carbon-Burnup Cell is to operate at 2,050°F. The deficit may be made up by feeding 10 pounds of coal per hour with the fly ash.

The sensitivity of this balance to some of the assumed parameters is as follows:

(1) If $\alpha$ = 0.8, the coal feed rate would have to be 250 lbs/hr.

If $\alpha$ = 1.0, the energy balance would indicate an energy surplus, so that 76 linear ft of 2" O.D. boiler tube would be added to the bed; or the air rate would be increased to about 1,260 lb/hr $ft^2$.

(2) If $A_e/A_p$ (See Equation 17) = 1.0; air rate would be increased to about 1,170 lbs/hr ft$^2$.

If $H_{eff}$ = 5 ft (See Equation 19), the coal feed rate = 425 lbs/hr or some refractory would replace the heat transfer surface on one wall.

## 6.6 Other Results from Tests in the FBC

### 6.6.1 Effect of Heat Removal

The effect of heat removal on combustion efficiency is shown in Figure 30.

The simple quadratic

$$CE = 100 \times (1 - 2.3\, X^2);\ \text{standard deviation}\quad S = 4.5\% \qquad (21)$$

was used between points 1 and 3 on the figure. Beyond point 3 the fire would go out if more heat were removed, so the curve must fall rapidly toward CE = 0.

The curve is felt to represent data for $T \simeq 1900$ to 2000°F, $A \simeq 600$ to 800 lbs/hr. For low air rates represented by point C, (A = 330 lbs/hr) an entirely different curve would apply.

Equation 21 is not useful as a prediction tool since the parameter "X" cannot be determined unless CE is already known, thus

$$X = \frac{\text{Energy lost to cooling surfaces and dust, Btu/hr}}{\text{Fuel rate, lbs/hr} \times \text{Carbon content} \times CE \times 14{,}100\ \text{Btu/hr}} \qquad (22)$$

The significance of Figure 30 can be understood by pointing out that the parameter "X" represents the fraction of the energy release not removed in the flue gas. So the

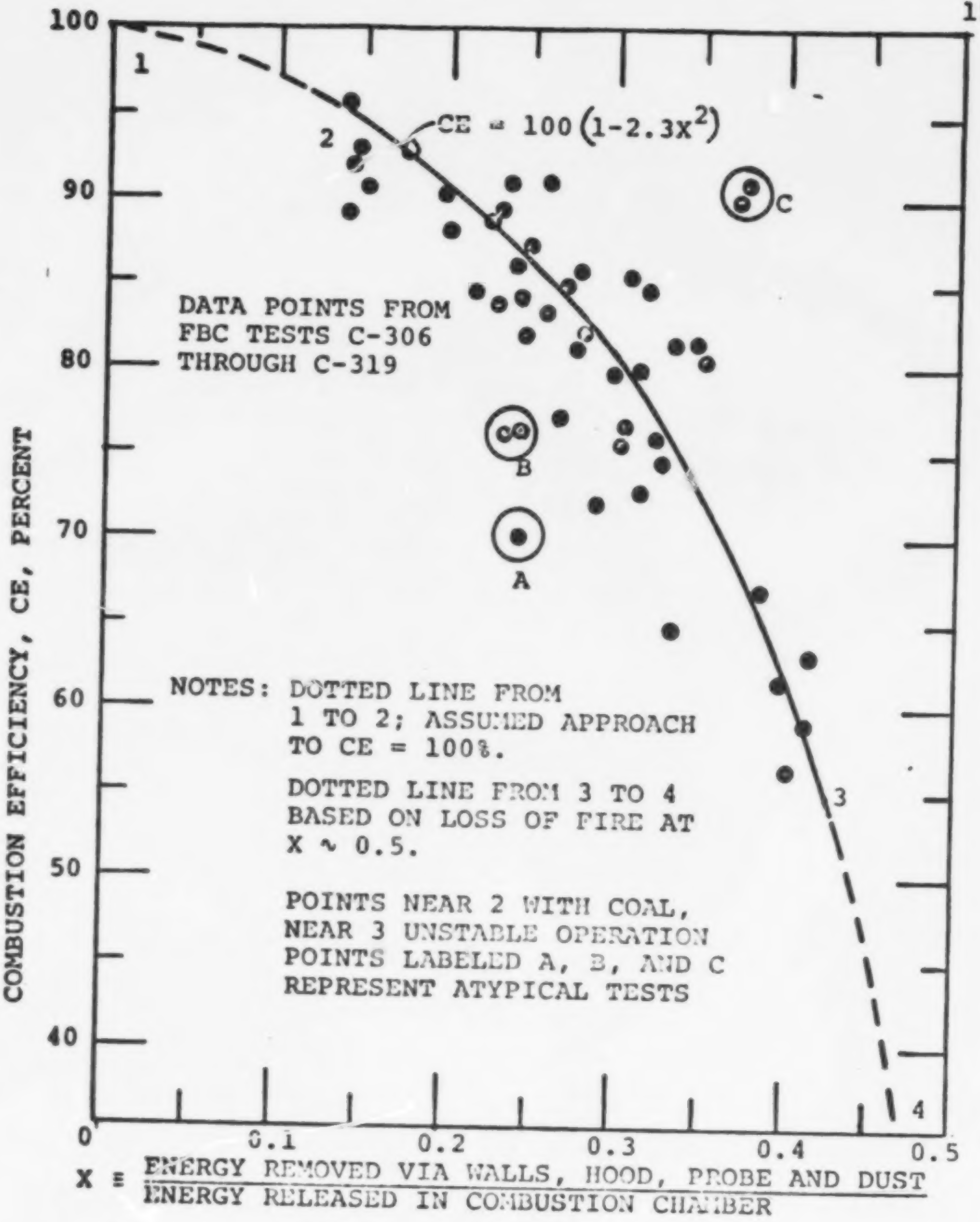


FIGURE 30 EFFECT OF HEAT REMOVAL ON COMBUSTION EFFICIENCY IN FBC

combustion should be most complete when only excess air is used for cooling. For the sample calculation in the preceding section the parameter $X \simeq 0.3$, and from Figure 30, $CE \simeq 81\%$ for this value of X, not the 90% used in the previous section. The model equations, are believed to be more reliable than the curve sketched on Figure 30.

### 6.6.2 Nitric Oxide Emissions

It has been shown in our past studies (1) that nitric oxide (NO) is the predominant form in the $NO_x$ group. This predominance was demonstrated by comparing IR measurements (specific to NO) with the PDS method which measures total $NO_x$. During the course of the current study, no attempt was made to measure total $NO_x$. It is assumed that the NO level recorded by the infrared analyzer continued to represent total $NO_x$. In past studies it was shown that when coal was burned in the fluidized bed, $NO_x$ increased with increasing residual oxygen in the range of 1 to 5% $RO(O_2)$. $NO_x$ showed no clear correlation with increasing bed temperature over the range $1500^oF$-$1850^oF$.

When the FBC was modified to simulate a Carbon-Burnup Cell and operated on fly ash in the range 3 to 10% RO and $1750^o$ to $2140^oF$, the dependence of $NO_x$ on residual oxygen was no longer apparent; in fact there was no correlation.

The statistical test series did not produce as useful a correlation between $NO_x$ and the control variables studied as for combustion efficiency, RO and $CO_2$. During the present tests, a general trend of increasing $NO_x$ with increasing bed temperature is shown in Figure 31. It may be seen that the highest values of NO were recorded at the highest temperatures, and the lowest values of NO were recorded at the lowest bed

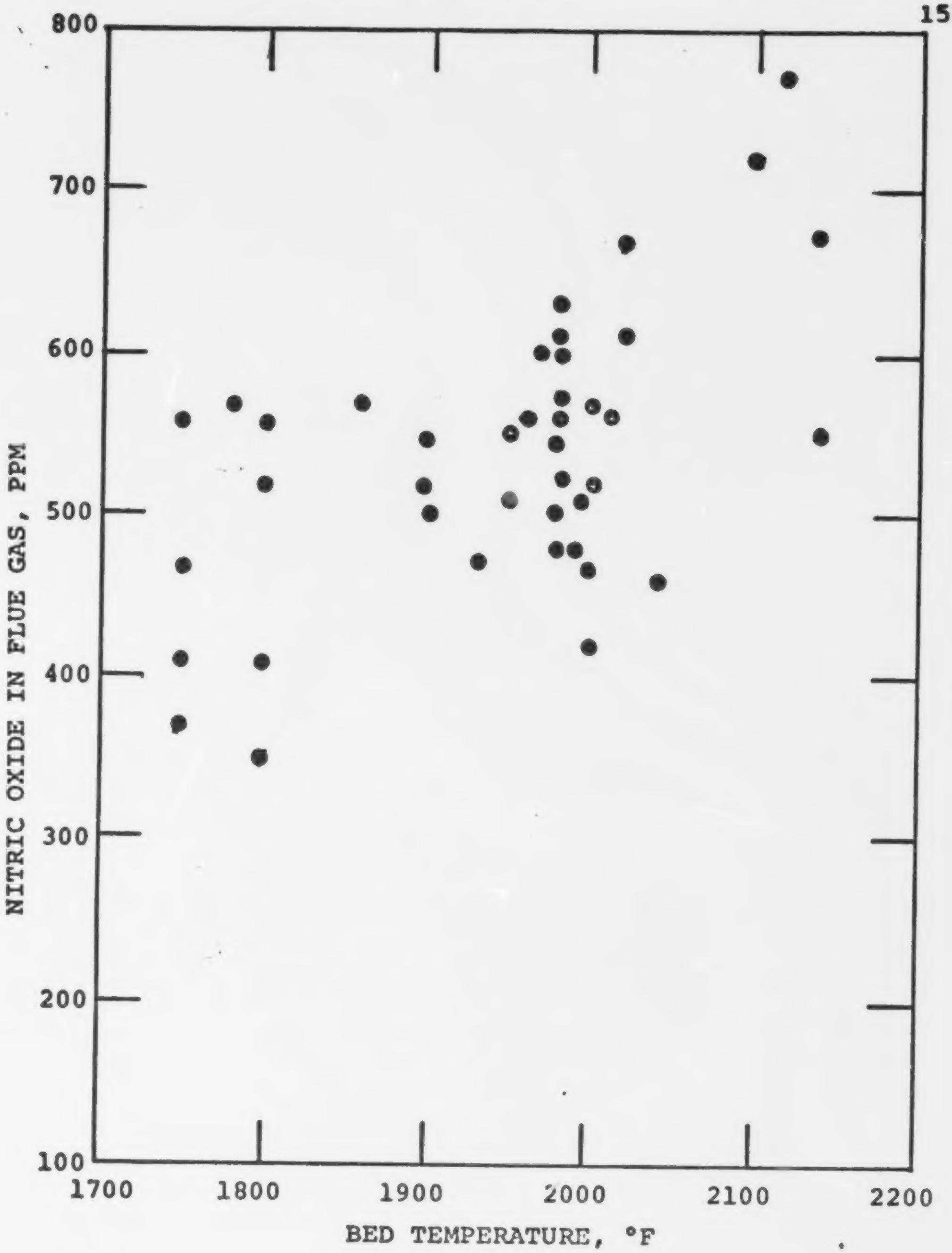


FIGURE 31 NITRIC OXIDE CONCENTRATION IN FLUE GAS FROM FBC VERSUS TEMPERATURE

temperature. However, the data are extremely scattered.

Although a clear correlation is not evident in Figure 31, an examination of the individual test results shows that within each test an increase in temperature almost always resulted in an increase in NO emission. Some typical results are plotted in Figure 32A.

When coal was added as in Run C-319, the NO level dropped with increasing coal content despite the high temperature and increasing residual oxygen (See Figure 32B). The results, in which coal and fly ash were interchanged, suggest that the major determinant in the level of NO emissions may be the combustion pattern within the bed, although a number of conflicting hypotheses have been advanced by other investigators. It is likely that NO was removed as formed, by reaction with hydrocarbons from coal.

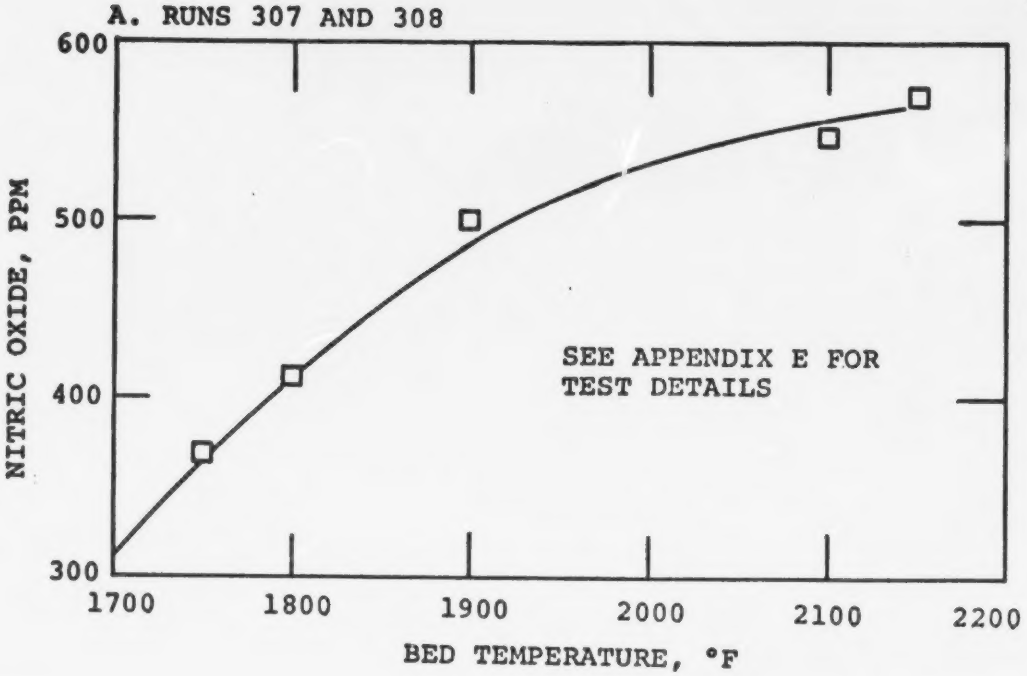
The correlation between the nitric oxide level and the control variables is as follows:

$$\begin{aligned}&\text{Nitric oxide, ppm} = -464.3\\&+ 0.4257\ (\text{bed temperature, }^{\circ}\text{F})\\&+ 0.329\ (\text{air rate, lb/hr ft}^2) + 4.468\ (\text{bed height, in.})\\&- 2.337\ (\text{carbon rate, lb/hr ft}^2) - 0.7829\ (\text{inert rate, lb/hr ft}^2) + 0.005372\ (\text{carbon rate x inert rate, lb}^2\text{/hr}^2\text{ ft}^4). \qquad (23)\end{aligned}$$

A comparison between the observed and calculated values is shown in Figure 33.

### 6.6.3 Sulfur Dioxide Emissions from FBC without Presence of Limestone Sorbents

Past studies (1) had shown that 90 to 95% of the



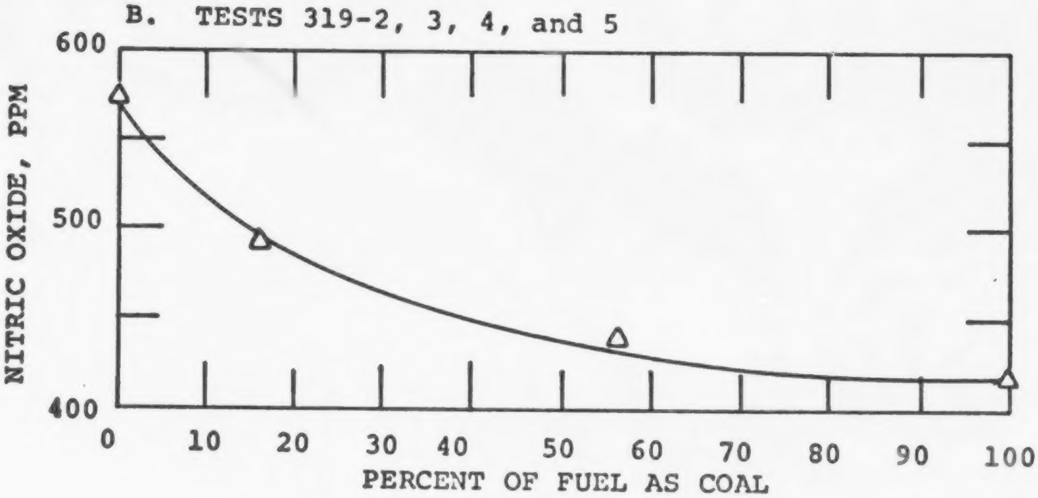


FIGURE 32 TYPICAL NITRIC OXIDE EMISSION RESPONSES IN FBC

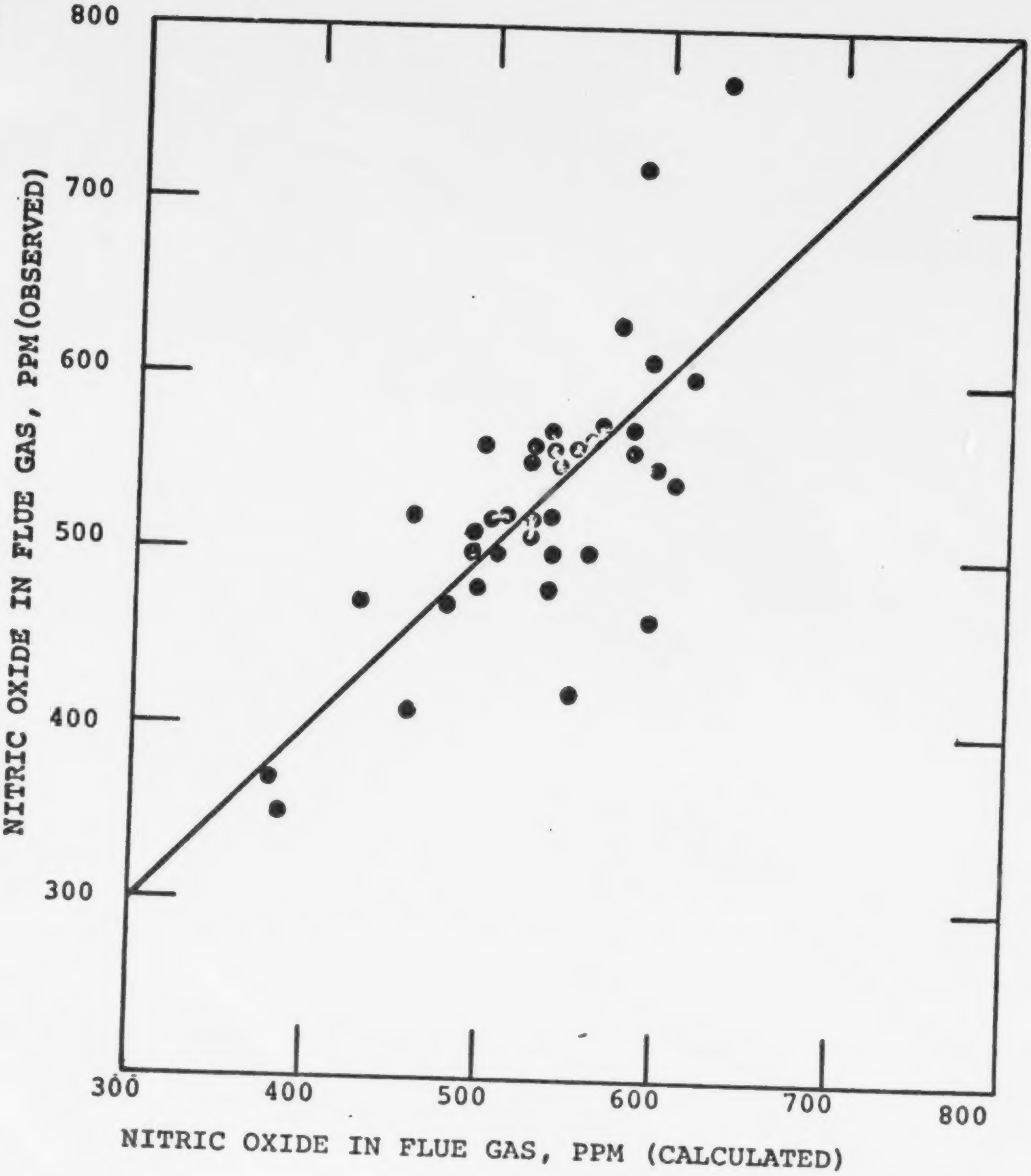


FIGURE 33 COMPARISON OF OBSERVED AND CALCULATED NITRIC OXIDE EMISSION FROM FBC

sulfur in coal would appear as $SO_2$ in the flue gas when the coal was burned in a fluidized-bed combustor without limestone addition. At the same time, a smaller fraction of the carbon would appear as carbon dioxide; typically, only 80 to 85% of the carbon was consumed. Fluidized-bed combustion of coal in a sorbent-free bed produces a fly ash which contains 5 to 10% of the sulfur and 15 to 20% of the carbon that was in the feed coal; hence, the requirement for a Carbon-Burnup Cell.

In burning the high-carbon fly ash during the present test program, the remaining sulfur was consumed. Table 17 below summarizes the sulfur balances for these tests and demonstrates that the sulfur and carbon in fly ash are consumed with about the same efficiency. Results for Runs 302 and 303 are presented, but are suspect and should be disregarded.

A ratio of combustion efficiencies defined by

$$\text{CE Ratio} = \frac{\%\ \text{sulfur burned}}{\%\ \text{carbon burned}} \qquad (24)$$

was computed for each test in Table 16. Excluding the results for Tests 302 and 303 the mean value was:

$$\overline{\text{CE Ratio}} = 1.0003$$

with standard deviation, s = 0.193

It is concluded that in the absence of a sulfur sorbent the sulfur and carbon are burned with equivalent efficiency in a Carbon-Burnup Cell.

### 6.6.4 Behavior of Limestone and Sulfur in a Carbon-Burnup Cell

In Section 6.4.6 the effect of limestone (in the

## TABLE 17

## SULFUR BALANCE SUMMARY FOR FBC TESTS

| Test No. | Fly Ash Sulfur Input lb/hr | Fly Ash Sulfur Output lb/hr | Flue Gas Sulfur Output lb/hr | Total Sulfur Output lb/hr | Input Less Output lb/hr | Sulfur Burned % | Carbon Burned % |
|---|---|---|---|---|---|---|---|
| C-302-1 | .605 | .248 | .286 | .534 | .071 | 47.2 | 90.0 |
| -2 | .860 | .310 | .640 | .950 | -.090 | 74.0 | 83.0 |
| -3 | .565 | .240 | .448 | .686 | .121 | 78.0 | 86.0 |
| C-303-1 | .410 | .310 | .110 | .420 | -.010 | 27.0 | 65.0 |
| -2 | .435 | .234 | .210 | .444 | -.011 | 48.3 | 66.0 |
| -3 | .615 | .402 | .180 | .582 | .033 | 29.4 | 65.0 |
| -4 | .578 | .320 | .266 | .586 | -.011 | 46.0 | 74.0 |
| -5 | .655 | .262 | .384 | .646 | -.009 | 58.7 | 78.0 |
| C-306-1 | .520 | .091 | .415 | .506 | .014 | 80.0 | 86.0 |
| -2 | .698 | .104 | .620 | .724 | -.026 | 89.0 | 70.0 |
| -3 | .635 | .082 | .530 | .612 | .023 | 83.0 | 89.0 |
| -4 | .535 | .070 | .450 | .520 | -.015 | 84.0 | 73.0 |
| C-307-1 | .510 | .194 | .306 | .500 | .010 | 60.0 | 76.0 |
| -2 | .292 | .097 | .210 | .307 | .015 | 72.0 | 90.0 |
| -3 | .518 | .070 | .420 | .490 | -.028 | 81.0 | 91.0 |
| -4 | .344 | .092 | .248 | .330 | .014 | 72.0 | 63.0 |
| C-308-1 | .470 | .085 | .398 | .483 | .013 | 85.0 | 81.5 |
| -2 | .550 | .086 | .450 | .536 | .014 | 81.0 | 81.3 |
| -3 | .560 | .075 | .470 | .545 | .015 | 84.0 | 80.6 |
| C-310-1 | .515 | .041 | .497 | .538 | .023 | 96.5 | 84.6 |
| -2 | .600 | .060 | .537 | .597 | .003 | 89.0 | 75.8 |
| C-311-1 | .600 | .130 | .448 | .578 | -.082 | 68.0 | 77.0 |
| -2 | .800 | .280 | .504 | .784 | -.016 | 63.0 | 66.7 |
| -3 | .950 | .350 | .525 | .875 | -.075 | 55.0 | 58.0 |
| -4 | .100 | .500 | .600 | .600 | .000 | 83.0 | 75.8 |
| -5 | .860 | .173 | NA | NA | --- | -- | -- |
| -6 | 1.11 | .275 | NA | NA | --- | -- | -- |
| C-312-1 | .645 | .135 | .530 | .665 | -.020 | 82.0 | 84.0 |
| -2 | .625 | .150 | .505 | .655 | -.030 | 80.0 | 83.8 |
| -3 | .670 | .175 | .510 | .685 | -.015 | 76.0 | 76.0 |

NOTE: In Tests C-302 and C-303 the fly ash feed was wet and not homogenous. The input data were probably not valid.

## TABLE 17

(Continued)

| Test No. | Fly Ash Sulfur Input lb/hr | Fly Ash Sulfur Output lb/hr | Flue Gas Sulfur Output lb/hr | Total Sulfur Output lb/hr | Input Less Output lb/hr | Sulfur Burned % | Carbon Burned % |
|---|---|---|---|---|---|---|---|
| C-313-1 | .87 | .08 | .76 | .84 | .03 | 87.3 | 91.0 |
| -2 | 1.03 | .10 | .87 | .97 | .05 | 85.0 | 87.2 |
| -3 | 1.17 | .21 | .98 | 1.19 | -.02 | 83.6 | 86.0 |
| C-314-1 | .90 | .22 | .67 | .89 | .01 | 74.5 | 84.8 |
| -2 | 1.01 | .23 | .76 | .99 | .02 | 75.0 | 82.0 |
| -3 | 1.33 | .29 | .90 | 1.19 | .14 | 68.0 | 76.0 |
| -4 | 1.16 | .14 | .98 | 1.12 | .04 | 84.0 | 89.5 |
| C-315-1 | .95 | .25 | .82 | 1.07 | -.12 | 86.2 | 83.3 |
| -2 | 1.01 | .20 | .89 | 1.09 | -.08 | 88.0 | 84.5 |
| -3 | 1.09 | .21 | .97 | 1.18 | -.09 | 89.0 | 85.2 |
| C-316-1 | .95 | .23 | .80 | 1.03 | -.08 | 84.5 | 74.5 |
| -2 | 1.11 | .23 | .95 | 1.18 | -.08 | 85.5 | 76.5 |
| -3 | 1.29 | .29 | 1.04 | 1.13 | -.04 | 80.5 | 72.0 |
| C-317-1 | 1.16 | .20 | .84 | 1.02 | .14 | 72.0 | 82.2 |
| -2 | 1.13 | .28 | .90 | 1.06 | .07 | 79.0 | 81.5 |
| C-318-1 | 1.04 | .35 | .78 | 1.13 | -.09 | 75.0 | 72.5 |
| -2 | .97 | .26 | .75 | 1.01 | -.04 | 77.5 | 79.8 |

fuel fed to a Carbon-Burnup Cell) on performance was outlined. It was concluded, on the basis of data presented there that:

(a) Limestone, or more precisely calcium oxide plus some sulfate, does not reduce the combustion efficiency of a Carbon-Burnup Cell.

(b) The performance model, Equation 6, was less accurate in predicting performance when the fly ash feed is rich in limestone.

(c) Under sufficiently oxidizing conditions, calcium sulfate will not decompose at the high temperatures experienced in a Carbon-Burnup Cell. This latter conclusion is discussed in more detail here and data presented on the conditions which lead to decomposition of calcium sulfate.

As noted in previous sections limestone may be fed, with the coal, to a fluidized-bed boiler in order to reduce air pollution due to sulfur oxides. Some, if not all, of the limestone would be carried out of the furnace by the flue gas and be removed by the dust collection equipment. The collected dust would contain unreacted carbon, ash which originated as mineral matter in the coal, and particles which contain calcium oxide, sulfate and possibly carbonate. Since no means exists for separating the carbon, the entire mass would be fed to the Carbon-Burnup Cell.

Some of the chemical reactions which could take place within the Carbon-Burnup Cell include:

$$2C(s) + O_2(g) \rightleftarrows 2CO(g) \qquad (25)$$

$$2CO(g) + O_2(g) \rightleftarrows 2CO_2(g) \qquad (26)$$

$$S(s) + O_2(g) \rightleftarrows SO_2(g) \qquad (27)$$

$$CaO(s) + SO_2(g) + 1/2\ O_2(g) \rightleftarrows CaSO_4(s) \qquad (28)$$

$$CO(g) + CaSO_4(s) \rightleftarrows CaO(s) + SO_2(g) + CO_2(g) \qquad (29)$$

The performance model, Equation 6, is a tool with which Equations 25 and 26 are linked to control variables. The material in the preceding section shows that, on the average, sulfur and carbon are burned equally well and so Equation 27 is linked to the control variables. Little data on Equation 28, the removal of $SO_2$ by lime, and Equation 29, the decomposition of calcium sulfate, were produced in the current test series. Data collected in earlier work (1) are of little value here since the conditions were so different in that work. Sufficient information was collected, however, to draw reasonable qualitative conclusions and to propose some interesting theories. These are discussed below.

### 6.6.4.1 Removal of $SO_2$ by Lime in a Carbon-Burnup Cell

In FBC Test C-318-2, the fly ash feed was precisely 1 wt. % sulfur. The fly ash contained no limestone. The ratio of sulfur to carbon was $\underset{-}{\sim}$ 1:60, by weight, while in the original #8 coal, fed to the FBM, the ratio had been $\underset{-}{\sim}$ 1:20. About 80% of the sulfur and carbon was burned in this FBC test ($T = 1950^{o}F$, $RO = 4.5\%$) and the $SO_2$ in the flue gas was 1050 ppm.

In FBM Test B-16, this same #8 coal was burned in the FBM and fine limestone injected. The composition of the fly ash which was then used in FBC Tests C-320-1 and 2, was given in Table 13. The ratio of sulfur to carbon in this material is $\underset{-}{\sim}$ 1:10, since much of the sulfur originally in the coal was present in the feed to the FBC as $CaSO_4$.

In Test C-320-1, almost 90% of the carbon was burned and it would be anticipated that 80 to 90% of the burnable sulfur would also burn. However, in this test the $SO_2$ in the FBC flue gas was only 350 ppm; or 1/3 the level in Test C-318-2.

The test conditions were comparable ($T = 1980^{o}F$, RO = 4.3%) to those in 318-2.

Two hypotheses are offered for the relatively low $SO_2$ level in Test 320-1:

1) Lime is capable of capturing $SO_2$ at temperatures approaching 2,000°F under the conditions of a Carbon-Burnup Cell.

2) When limestone is added to a fluidized-bed burning coal, the limestone acts to reduce the quantity of sulfur in the fly ash which is subject to oxidation; i.e., limestone converts fly ash sulfur into calcium sulfate.

The first hypothesis is supported by the findings of many investigators that the kinetics of $SO_2$ capture by lime are enhanced by increasing temperature in conventional furnaces and with artificial flue gas. But, these results are in direct contradiction to our earlier work (1) and that of almost every other worker (23) in fluidized-bed combustion.

One major difference between a Carbon-Burnup Cell and a coal-burning fluidized-bed combustor is that, in the former, little hydrogen is contained in the fuel; i.e., hydrogen or methane from the volatiles in coal might reduce the efficiency of sulfur capture in the FBM by:

$$4\ CaSO_4\ (s) + CH_4(g) \rightleftarrows 4\ CaO(s) + CO_2(g) + 2\ H_2O(g) + 4\ SO_2(g) \qquad (30)$$

Reaction 30 is a global one and does not specify whether the intermediate calcium - sulfur compound is sulfide or sulfite. If sulfide, the intermediate could be oxidized by oxygen in the gas or by adjacent molecules of sulfate in the solid. A debate on mechanisms is not useful at this point.

The second hypothesis is simply a conjecture that the history of sulfur in coal might be as follows:

Assume that most, or even all, of the sulfur in coal leaves the coal particle as it is being pyrolized, and that the resulting sulfurous gas may attack lime or residual carbon (or oxygen and so form $SO_2$). In the absence of lime, some of the sulfur recombines with the carbon to appear in the fly ash as a residual organic sulfur. When this fly ash is subsequently burned, this organic sulfur is oxidized and appears as $SO_2$ in the flue gas.

If lime is present, in sufficient excess when the coal is being pyrolized, sulfur is fixed by the lime rather than by carbon and enters the fly ash stream as $CaSO_4$, not $C_xS_y$. When this fly ash is subsequently burned, the sulfur, as $CaSO_4$, remains in that form and does not appear as $SO_2$ in the flue gas.

The obvious experiments and analyses which might have proven which hypothesis was correct, if either really was, were not done; the program was essentially over and the reason didn't matter -- it had been definitely established that $SO_2$ emissions from a Carbon-Burnup Cell would be low if limestone were fed to the primary cells, and the burnup cell were designed and operated properly.

### 6.6.4.2 Release of $SO_2$ by Lime

Equations 28 and 30 explain, in a straightforward manner how the sulfur in lime can be released as $SO_2$. Other paths are as follows:

$$CaSO_4(s) + CO(g) \rightleftarrows CaSO_3(s) + CO_2(g) \quad (31)$$

$$CaSO_3(s) \rightleftarrows CaO(s) + SO_2(g) \quad \rightleftarrows \text{(at } T > 1300^{\circ}F) \quad (32)$$

or

$$4CaSO_3(s) \rightleftarrows CaS(s) + 3CaSO_4(s) \quad (33)$$

$$CaS(s) + 3CaSO_4(s) \rightleftarrows 4CaO(s) + 4SO_2(g) \quad (34)$$

A re-examination of previous $SO_2$ control results (1) was made in an effort to explain the release of $SO_2$ by lime, and these results were compared with predictions based on published thermodynamic data. These published data are highly suspect, since data on calcium compound thermodynamics have never been rigorously reviewed and do not appear in the JANAF Tables.

The difference between predicted and observed results led to the postulation of a pseudosulfate as an intermediate between $SO_2$ capture and the $CaSO_4$ product. The sulfur bound as the pseudosulfate was assumed less tightly held than in the true sulfate, and would decompose more readily than the true final sulfate. Appendix H contains the theoretical analysis and the results of an experiment intended to determine if psuedosulfate played a major role in the release of sulfur from partially sulfated lime.

The experiment, C-320-2 and C-320-3, was carried out by firing, in the FBC, fly ash from the FBM containing the lime by-products and then switching to an artificial mixture of coal fines, ash, limestone, and natural gypsum. The composition of the natural fly ash is described in Table 13. The test revealed that both feed materials produce increasing $SO_2$ as residual oxygen is reduced (See Figure H-2), and that the pseudosulfate theory was not necessary to explain the observed results. It was noted that it is possible to retain the sulfur in the artificial mix and in the FBM fly ash by operating the FBC at residual oxygen levels above about 3%.

Table 18 shows the level of $SO_2$ in the FBC off-gas as a function of residual oxygen with a fly ash/lime by-product feed.

TABLE 18

$SO_2$ APPEARING IN EXHAUST OF FBC FIRING FLY ASH CONTAINING 19.4% Ca, 2.4% S

Bed temperature = 1970-1980°F, Air rate = 620 lb/hr
Fuel Rate = 167-174 lb/hr, Superficial velocity = 12.1 fps

| Residual $O_2$, vol. % | $SO_2$, ppm |
|---|---|
| 0.1 | 4100 |
| 0.4 | 2400 |
| 0.5 | 2150 |
| 0.8 | 750 |
| 1.0 | 550 |
| 4.3 | 350 |

It may be seen from these results that fine lime, in fly ash, cannot be effectively regenerated under the conditions existing in a Carbon-Burnup Cell.

1) The residence time of the fine particles is apparently too short to permit more than partial desulfurization. or

2) $SO_2$ may be re-absorbed in the flue as the gas and lime particles travel toward the dust collector together.

Whatever the reason, only about one-half the calcium sulfate was decomposing at 0.1% $O_2$.

### 6.6.5 Particulate Emissions

The fuel fed to a Carbon-Burnup Cell will be relatively high in ash. Without the addition of limestone, the fuel would typically be 50% ash or about 35 pounds of ash per $10^6$ Btu fired. With limestone, ash (or inert) input would be on the order of 240 lbs per $10^6$ Btu fired. These values may be compared with a typical coal (9% ash, 13,500 Btu/lb) where the ash input is about 7 lbs per $10^6$ Btu fired.

In tests in the FBC the particulate emissions, PE -- that is the fly ash passing through the bed as well as the FBC's cyclone collector -- were

$1 \lesssim PE' \lesssim 8$ lbs per million Btu of fuel fired.

The emission data, in this form, were plotted against reciprocal bed temperature in Figure 34. These data were too scattered to produce a useful correlation, and none should be expected considering the number of control variables. (These are all considered later). It is clear however that emissions decline with increasing bed temperature.

The reduction in emissions with increasing bed temperature is expected because:

a) At high temperature more of the carbon burns and so less is available as particulate. This has been demonstrated previously.

b) At high temperature, some of the ash matter is retained by the bed particles via agglomeration. Also, agglomeration makes those particles which leave the bed larger and hence more collectable by the cyclone. These factors are discussed below.

Goldberger (17, 23), burning coal in a fluidized-bed of sand, has demonstrated that some of the ash would be retained by the bed particles according to:

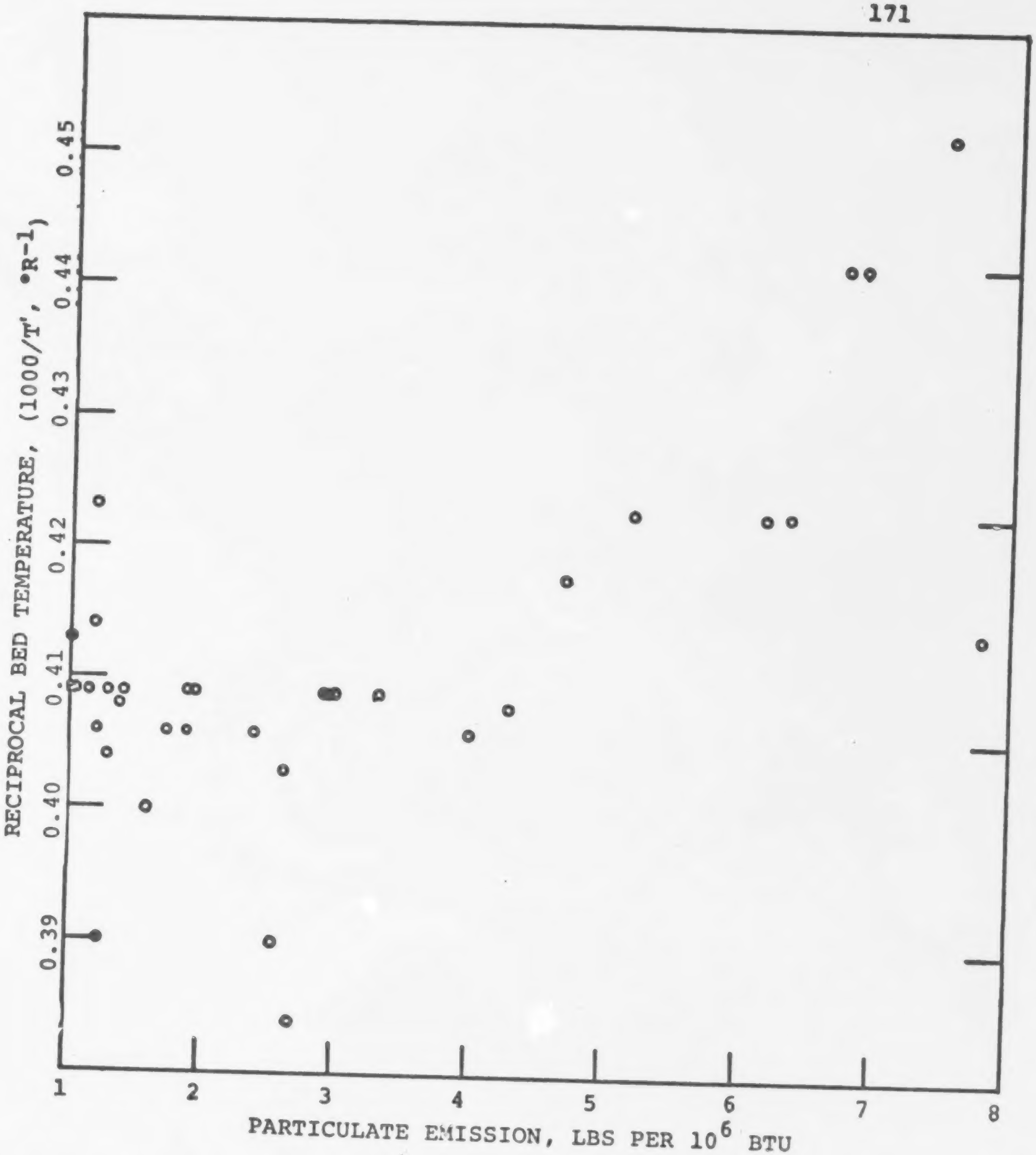


**FIGURE 34** PARTICULATE EMISSION FROM FBC AS A FUNCTION OF BED TEMPERATURE

$$\ln \frac{C_{in}}{C_{out}} = K_c \frac{W}{G} \qquad (35)$$

where

$C_{in}$, $C_{out}$ = fly ash concentration entering the system and leaving the fluidized-bed, grains/scf

$k_c$ = capture coefficient, grains captured/min x lb-bed x grains per scf

W = bed weight per unit cross section, lb-bed/ft$^2$

G = gas flow rate per unit cross section, scfm/ft$^2$

Goldberger showed that

$$\ln k_c \simeq -51{,}300 \left(\frac{1}{T'}\right) + 20.47 \qquad (36)$$

where

T' = bed temperature, $^{o}$F

From Equations 35 and 36 it follows, that the quantity of dust leaving the bed and entering the dust collector will decline as the:

1) quantity of entering mineral matter is reduced
2) bed temperature increases
3) bed depth (or particle density) increases
4) air rate is reduced

Also, the fraction of dust entering a cyclone dust collector which passes through without capture will decline as the:

1) quantity of entering dust is increased

2) bed temperature is increased (because this acts to increase the gas temperature, hence velocity entering the

collector)

3) air rate is increased

Goldberger finally suggests that

$$E_s, \% = \frac{\text{total ash collected by system}}{\text{total ash in feed}} \times 100 \qquad (37)$$

would be given by

$$E_s, \% = 100 -(\exp \ -k_c(W/G) \times (100-E_c) \qquad (38)$$

where

$E_c$ = dust collection efficiency of cyclone, % (considered independent of T' and G)

Attempts to reproduce the experimental data derived from the tests in the FBC with Equations 35 and 36 were not successful. This may be because Golderger's studies were carried out in the range $1 \lesssim W/G \lesssim 3.5$, while most of PER's work was carried out with $W/G \lesssim 1$.

A second reason may be that the ash matter in fly ash has a lower propensity to agglomerate than the ash in coal.

However, there was some indication that at the highest temperatures the ash was captured. For example, during tests C-302-2 and C-306-3, where the bed temperatures exceeded 2,100$^{\circ}$F, the bed mass was increasing so rapidly that material had to be removed to maintain a constant reading on the low bed manometer which indicates bed depth. The amount of material removed was not noted.

The agglomeration effect at high bed temperature may also be noted in Figure 35. This figure presents the results of a sieve analysis of bed samples taken before and after run C-308, in which the bed spent a short time (about 15 minutes) at 2,100$^{\circ}$F and about two hours at 1,900$^{\circ}$F. The original top

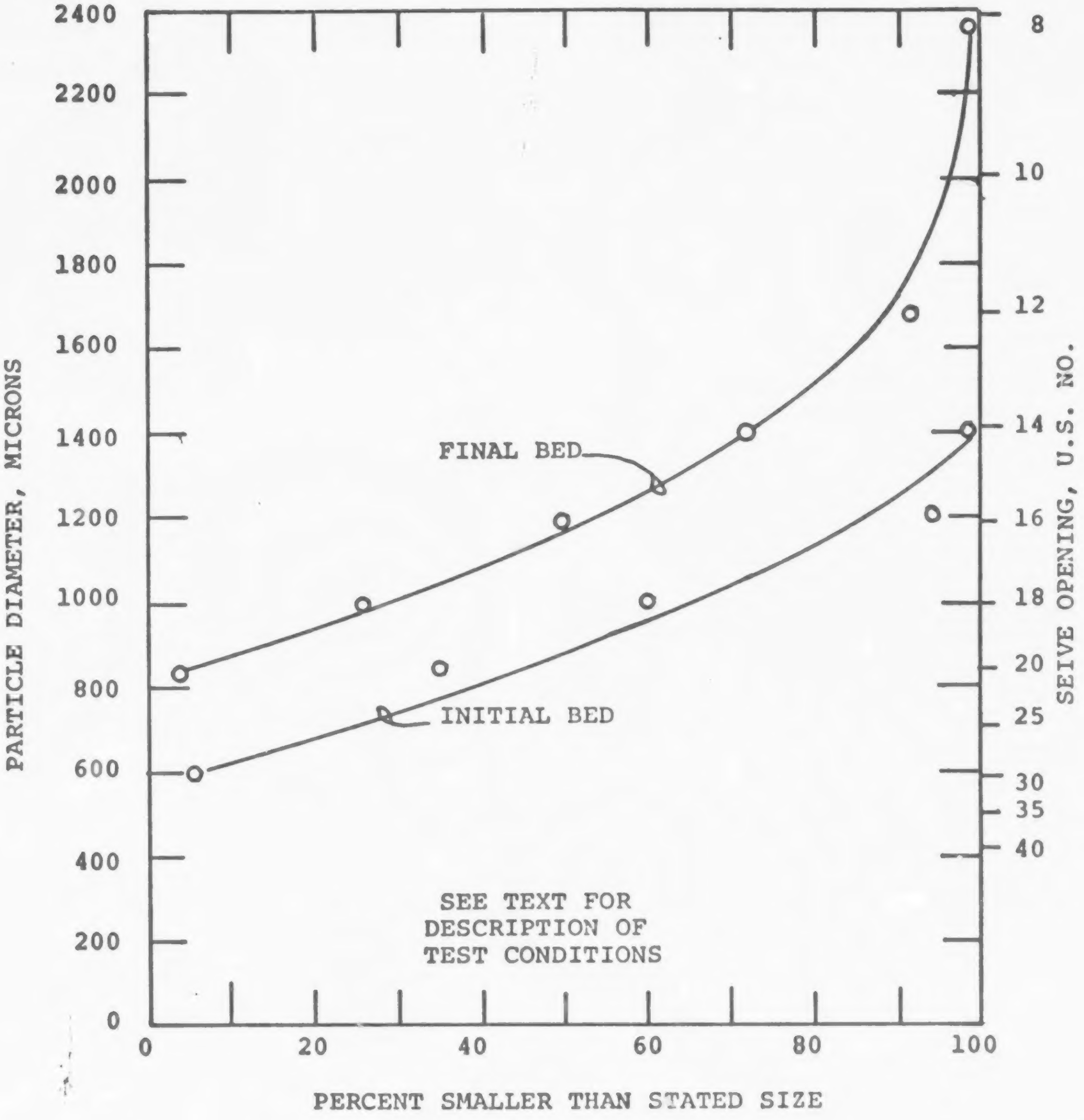


FIGURE 35 ASH BED PARTICLE SIZE DISTRIBUTION BEFORE AND AFTER HIGH TEMPERATURE OPERATION IN FBC TEST C-308

size before the run was 14 mesh. It is seen that 30% of the bed particles after the run were larger than the original top size. Excluding a sampling error, this result can only be explained on the basis of agglomeration.

a) The elutriation of bed fines could not have increased the top size

b) The fly ash fed had a top size less than 1000 $\mu$m

Also, when the bed particles were examined they had a coating of ash matter.

It is clear that the ability to agglomerate fly ash in a Carbon-Burnup Cell may be a marked advantage of the Multicell Fluidized-Bed Boiler.

In certain tests where limestone was present (to be discussed in a future report), fly ash formed uniform spheres, up to about 1/2" in diameter. At the least, ash in this form could be disposed of at no cost to the boiler operator since it would be a useful fill material.

The particulate emission data listed in Appendix E was fit to the performance model (See pages F-11 through F-13). Equation 3, based again on the assumption that cross sectional area is the appropriate scaling factor, was the result.

$$\begin{aligned} PE = {} & \text{Pounds of particulate emission per hour per} \\ & \text{ft}^2 \text{ of bed cross section} = \\ & + 15.57 - 0.00664(T) - 0.00134(A') \\ & + 0.0236(H) - 0.03034(C') + 0.00461(I') \\ & + 0.00034(c' \times I') \qquad (3A) \end{aligned}$$

The signs of the regression coefficients and their relative importance appear logical, except for that modifying

H, the static bed depth. It would be expected that coefficient would be negative. This is not a serious flaw since the magnitude of the entire term is low.

The derivation of this relationship is discussed in Appendix F. The calculated and observed data are compared in Table F-4.

### 6.6.6 Hydrocarbon and Carbon Monoxide Emissions

Hydrocarbons were measured continuously by a flame ionization analyzer, while carbon monoxide was measured intermittently by Orsat analysis. During the entire FBC test series, the highest hydrocarbon value was 200 ppm, at a time when coal fines were used with gypsum to test the pseudosulfate hypothesis at a residual oxygen level of 0.8%. The next highest value was 50 ppm, and most readings (reported in Appendix E) were 30 ppm or less. Orsat CO readings were zero. However, one scale division on the apparatus is 0.2%. Therefore, under carbon burnup cell conditions (high temperature and high residual oxygen, without coal feed), little if any hydrocarbon or CO emission is anticipated.

# 7.0 DEVELOPMENT OF AN INTEGRATED FLUIDIZED-BED BOILER AND CARBON-BURNUP CELL: TESTS WITH THE FBM/CBC

## 7.1 General

As described in Section 5.2.2, a simulated Carbon-Burnup Cell was added to the modular boiler cell (FBM). As was noted, this unit, designated the CBC, was added to evaluate the Carbon-Burnup Cell concept in an integrated system and to gain insights on the operation of a Multicell Fluidized-Bed Boiler.

The CBC's air distributor had the dimensions 10-1/2" x 15-1/2". Compared with the FBC, the CBC was 28% larger in plan area. Unfortunately, the effective height for disentrainment of particles splashed above the bed of the CBC was considerably less than that of the FBC. This was caused by the existing FBM steam drum arrangement.

The FBM and CBC shared a common fluidized bed, in that an opening existed in the common wall shared by the two units. Bed particles would therefore circulate between the two reactors at a rate established by a diffusion-like driving force. The interchange was sufficiently rapid to permit the CBC to be ignited by the hot bed material which flowed in from the operating FBM, without the use of a separate gas-fired ignition system. This feature, and several others, characterize the Multicell Fluidized-Bed Boiler.

If the exchange of bed material were too rapid, the desired temperature differential (300°F to 600°F) required for efficient operation could not be maintained One purpose of the test effort with the FBM was to determine what the interchange rate was and how it was controlled.

In adding the CBC to the FBM, it was necessary that the coal and additive injection systems be changed from those described in Reference 1. A technique of vertical coal injection was seen as the preferred mode for commercial application; so, coal formerly injected via the front of the unit at operating floor level was now to be injected via the top of the unit. This also cleared the operating level for the equipment required for the CBC.

The coal injection method and the efficiency of sulfur capture by limestone are related through the oxygen requirement of reaction 28 and the deleterious effect of reducing gases implied by reactions 30, 31, and 32. If the quality of the coal distribution is poor, regions will exist in the combustion bed in which there is insufficient oxygen to carry reaction 28 to the right; i.e., to $CaSO_4$, and $SO_2$ will escape from the bed into the flue gas. If grossly reducing conditions were to exist, $CaSO_4$ might tend to decompose and $SO_2$ escape the bed into the flue gas. Therefore, another purpose of the test program was to determine the effectiveness of coal distribution in a large fluidized-bed combustor.

Because of the limited space above the bed in the CBC, particles of bed material which were thrown into the freeboard were swept out with the exhaust gas and appeared in the unit's dust collector. Although this effect was very undesirable, compactness was seen as a desirable feature of a fluidized-bed boiler. Therefore, another purpose of the test effort was to determine whether bed material loss could be inhibited by a baffle screen, and how effective heat exchange would be in the freeboard of a fluidized-bed combustor.

Fine limestone injection had been studied in the program reported in Reference 1. The addition of the CBC to the FBM and the changes in the coal and additive injection systems required that the efficiency of fine lime injection be re-evaluated. There was some concern that the high temperatures of the CBC would cause the calcium sulfate formed in the FBM to decompose in the CBC. This effect was potentially advantageous--if the fine lime could be reused in the FBM or, for example, if Portland Cement and sulfuric acid were to be produced from coal ash and partially sulfated lime. More likely, decomposition of fine limestone by-products at the conditions of a Carbon-Burnup Cell could not be exploited. Therefore, another purpose of the FBM test program was to assess fine limestone effectiveness in an integrated system and to determine the fate of sulfur carried into the CBC with the fly ash.

## 7.2 Coal Feeding to the FBM

### 7.2.1 General

As noted earlier, coal had previously been fed into the FBM via a horizontal pneumatic injector which rested on the air distribution grid. This method is described in detail in Reference 1. Other coal injection methods tested are described in Reference 4. As part of the current program, it was necessary to revise the coal injection technique to accommodate the addition of the CBC to the FBM. This provided the opportunity to test a vertical coal injection technique, which was felt to have a number of potential advantages. The revised coal handling system was shown schematically in Figure 10. The portion of the injection system internal to the boiler was shown in Figures 8 and 9. It was this internal portion that most directly influenced performance.

Briefly, the external system was as follows: Coal is carried from the weigh hopper to the pneumatic injector via an inclined screw. This inclined screw provides an effective pressure seal between the 35" w.g. pressure at the injector inlet and the atmospheric pressure above the open weigh hopper. The coal dropping off the end of the screw passes through a flexible connection (required by the weighing arrangement) and enters the vertical pneumatic injector.

The vertical injector consists of a section of 1-1/4" square tubing of 304 stainless steel which penetrates the upper boiler wall just below the steam drum. The feeder divides into two parallel tubes, also 1-1/4" square, which go down through the fluid bed and which terminate with a horizontal turn, one feeder pointing forward and the other toward the back of the FBM. The feeder assembly is centered in the boiler and the coal is discharged about 1" above the grid plate. Each outlet tube services about 4 square feet of grid area. As described in Reference 4, a single point injector servicing 8 square feet would be less effective.

### 7.2.2 Results

For a fluidized-bed boiler, the quality of fuel injection is determined by the presence or absence of temperature gradients within the heavily cooled fluidized bed, and by mapping residual oxygen distribution above the bed. Symptoms of poor fuel distribution would be: large horizontal temperature gradients, large horizontal oxygen, CO and hydrocarbon gradients, poor sulfur capture by limestone, and under some operating conditions, the agglomeration of large quantities of bed material. In early work described in Reference 4, horizontal gradients of 250°F over a 6 foot span were not unusual.

Poor fuel distribution within the bed may result from inherent flaws or design errors. As an inherent flaw we might include, for example, feeding gas, oil, solid waste, or pulverized coal into the bed of a fluidized-bed boiler through point injectors. As design errors, we might include: arranging the heat-exchange surfaces so that they interfere with the movement of coal through the bed, or providing an inadequate number of discrete coal injection points; i.e., over-estimating the mixing ability of the fluid bed or under-estimating the reactivity of the fuel. Several other potential design errors are discussed below.

The initial shakedown test of the modified FBM system gave evidence of extremely poor coal distribution. In this test, Feeder "a" in Figure 36 was used. Over the 6-foot span of the unit, residual oxygen, measured 86" above grid level,* was negligible directly above the coal injector and over 10% at the two ends of the unit. (10% is the limit of the recorder). Investigation revealed that the injector had partially plugged with coke, a symptom of the coal overheating in its passage through the feeder, possibly from inadequate injection velocity, or from excessive residence time. For transport air at ambient temperature, the temperature rise of the coal traveling through the feeder had been calculated to be less than 300°F. This calculation indicated that the coal would not be overheated--that no devolatilization or coking would take place. The original 1" x 2" injector was replaced with the 1-1/4" injector to increase velocity, and the tube above the bed wrapped with insulation to reduce heat gain from the bed and freeboard. This modification is shown as "b" in Figure 36. The symptoms of poor coal distribution; i.e., horizontal oxygen gradients, continued during subsequent shakedown tests. A water jacket was

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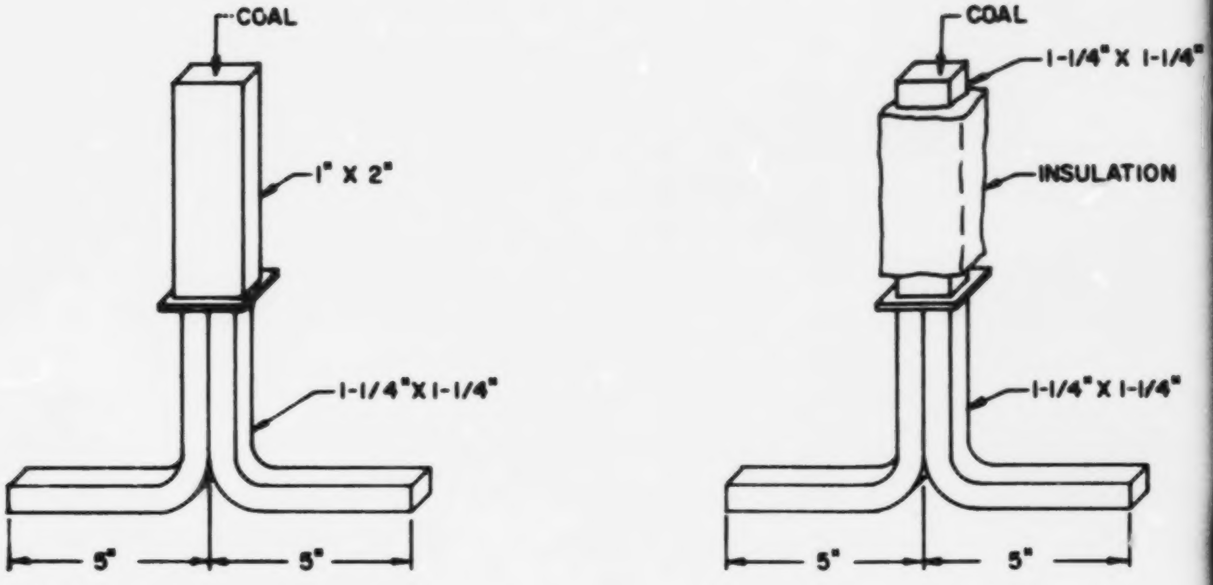
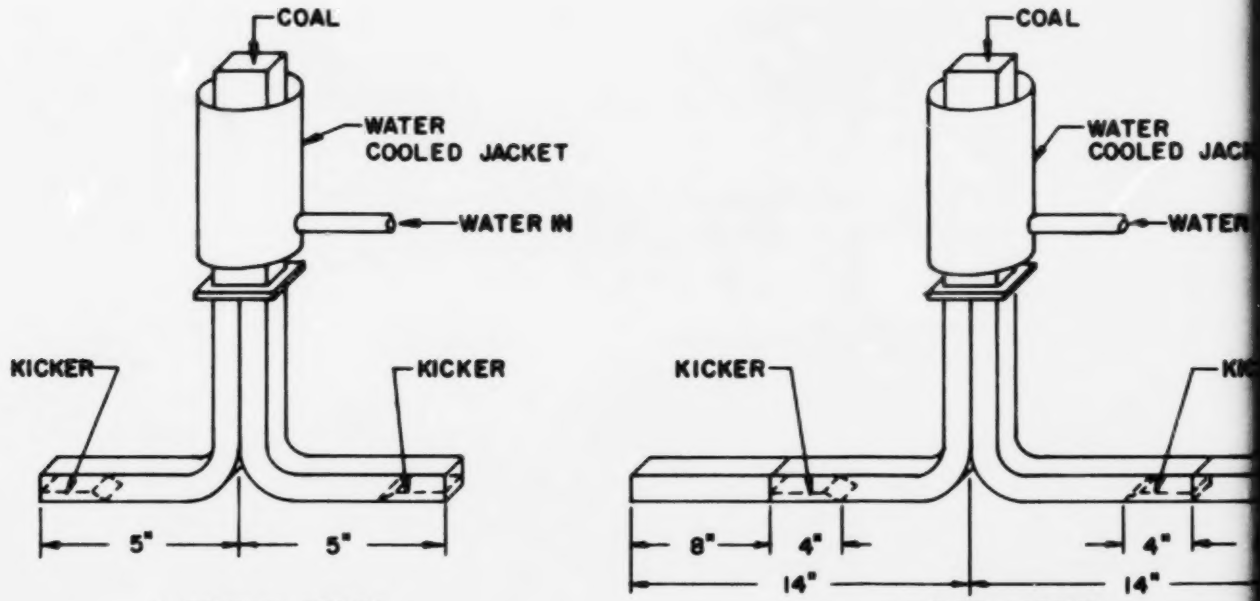
*In more recent tests a probe 4 ft. above grid was used.



a. FIRST DESIGN

b. SECOND DESIGN



d. FOURTH DESIGN

e. FIFTH DESIGN

FIGURE 36 FBM COAL INJECTOR MO

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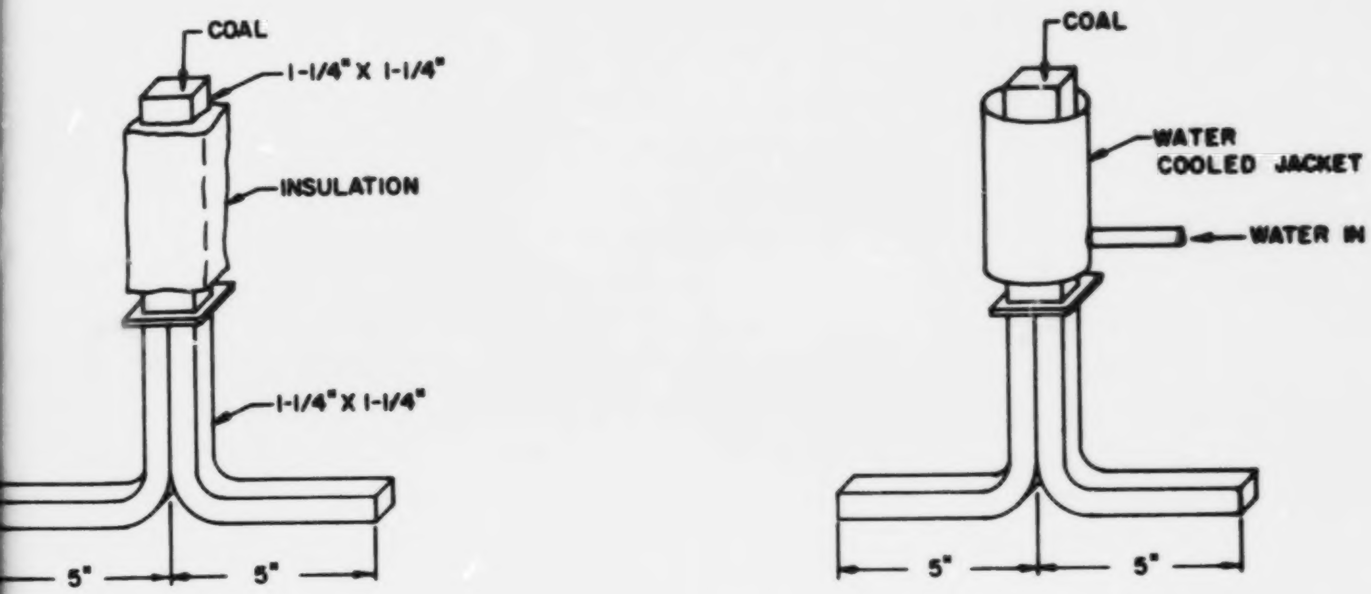
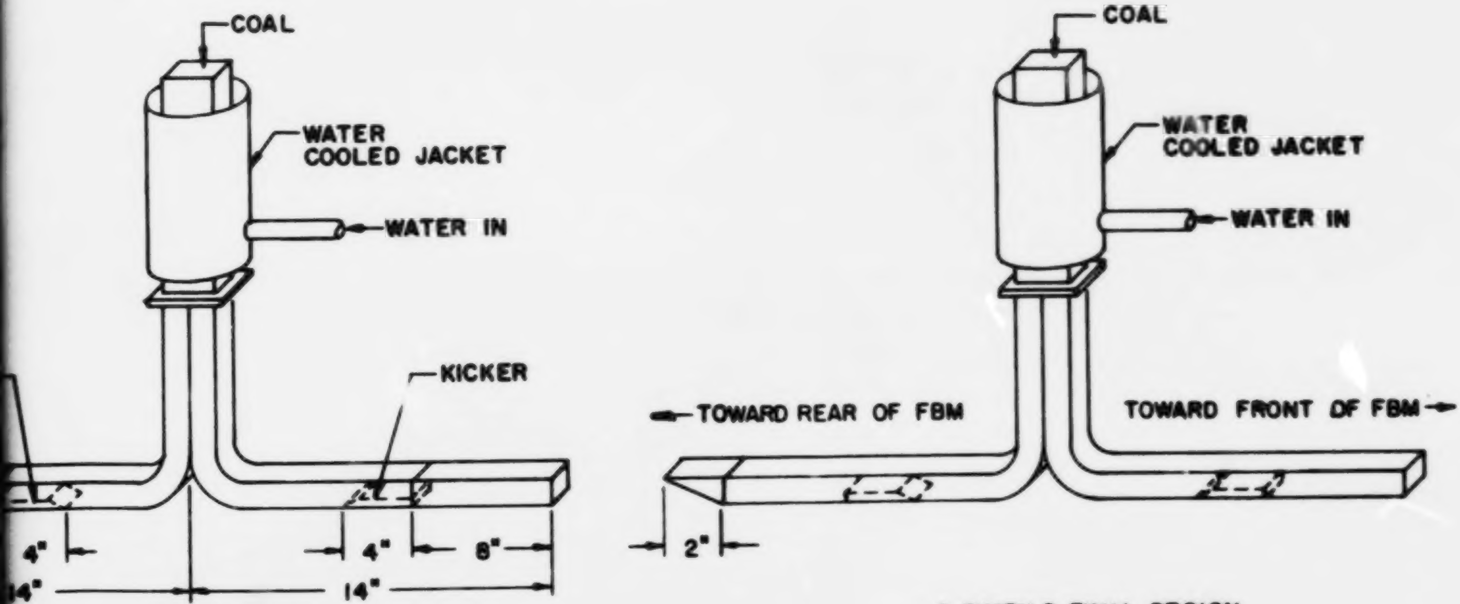


b. SECOND DESIGN

c. THIRD DESIGN



e. FIFTH DESIGN

f. SIXTH & FINAL DESIGN

FBM COAL INJECTOR MODIFICATIONS

POPE, EVANS AND ROBBINS
CORPORATED

B

added to the coal injector as shown in "c" of Figure 36, and still no improvement was noted in the oxygen map. It was then apparent that the calculation of coal temperature rise had probably been correct and that the poor coal distribution was the result of some other, as yet unknown, design error.

It was suggested that the poor performance of this new vertical injector could be explained if the larger coal particles were being disentrained in the 90° bend between the vertical and horizontal sections of the injector. Once the large coal particles had been separated from the injection air stream, they would not enter the fluidized bed with a velocity near that of the injection air. Instead, it was visualized that the injection air, and fine coal left the feeder at the top of the tube at high velocity, while the coarse fraction would "slide" out of the bottom of the tube without the momentum required to deeply penetrate the fluidized bed. To cause the coarse coal to be re-entrained, a "kicker" plate was tack-welded into the horizontal section, as shown in "d" of Figure 36.

The addition of this kicker was not a complete soltition, but the extreme oxygen gradient previously encountered was reduced to a 5% $O_2$ differential between the center of the boiler and either end. A typical set of $O_2$ readings here would be a 1% $O_2$, at center, 6% at either end. The average gas composition, measured downstream, would then range from 3.5 to 4.5%.

Extension of the horizontal legs to 14", as in "e" of Figure 36, gave a favorable oxygen map. The typical maps with this feeder configuration is shown in Figure 37 for FBM Test B-3. $\Delta O_2$ is about 1 percentage point. Figure 37 shows that the fuel/air ratio was slightly higher in the rear of the

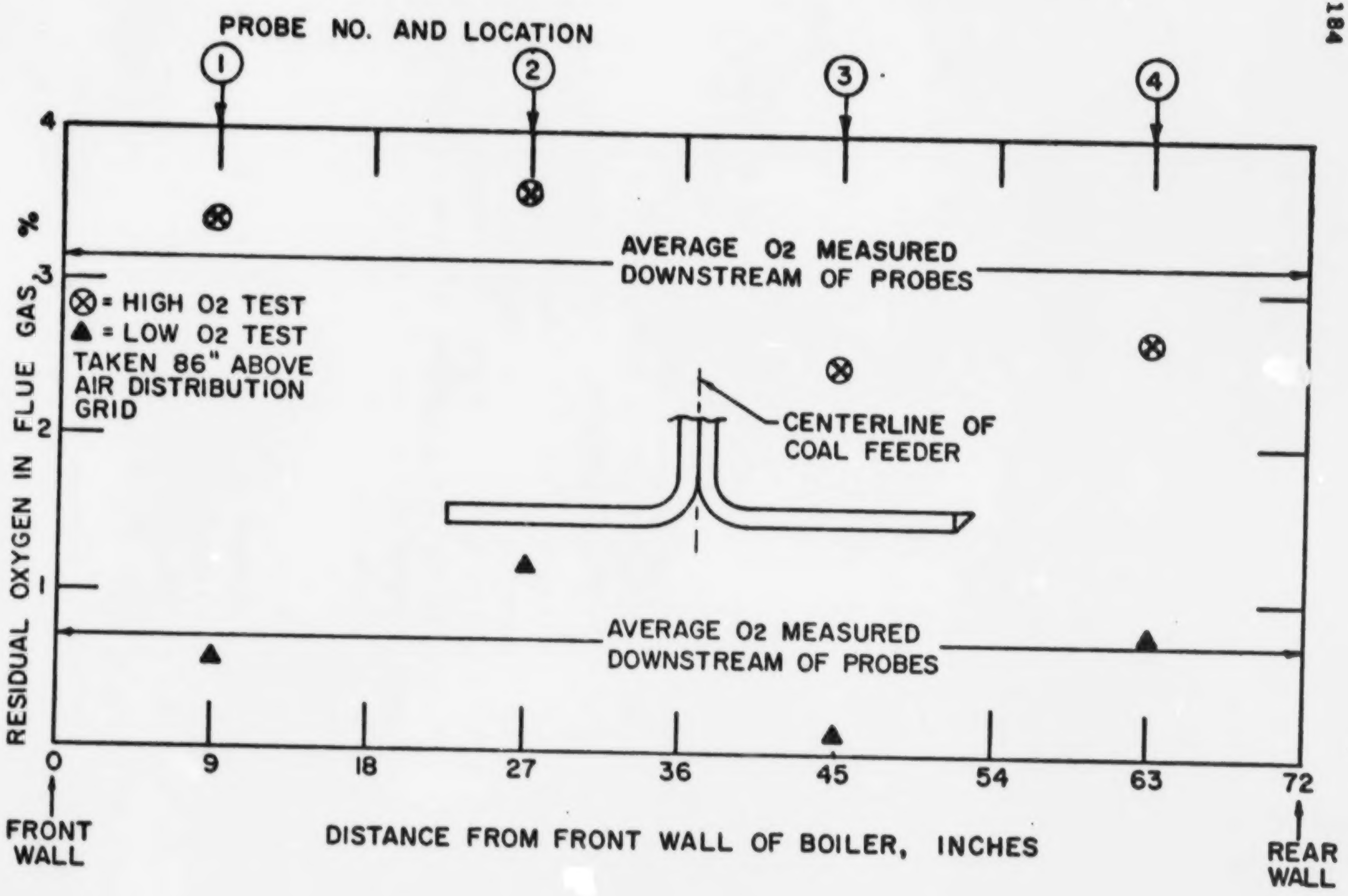


FIGURE 37 TYPICAL FBM RESIDUAL OXYGEN MAPS- FBM TEST B-3

unit. With the feeder further modified as shown in "f" of Figure 36, the oxygen gradients were reduced to as low as ±0.2% $O_2$ over the 6-foot span.

## 7.3 Feeding Fly Ash to the CBC

### 7.3.1 General

The use of fine solids ( < 20 mesh) in a fluidized bed, is not unusual. Most commercial fluid-bed processes, in fact, make use of fine solids, and in order to avoid excessive loss of these particles by elutriation, limit superficial gas velocities to low levels. For particles the size of fly ash, about 100 microns, the terminal velocity, the highest velocity at which particles would remain fluidized and not elutriate, is on the order of one foot per second.

Operation of a Carbon-Burnup Cell at air rates that would result in a superficial velocity of one foot per second was undesirable for two reasons. First, it was necessary to build the system compactly and this requires that relatively high air rates be used. Second, for a Carbon-Burnup Cell which is an integral part of a fluidized-bed boiler, the minimum permissible superficial velocity, set by the size and density of the bed particles in the coal-burning regions, will normally be far in excess of one foot per second. But high air rates are incompatible with long residence times for fine particles in the fluidized bed of the Carbon-Burnup Cell.

If the fine particle residence time is short, the time available for reaction within the bed will also be short. For a bed one foot deep, operating at an air rate and temperature which yield a superficial gas velocity of 10 feet per second, the average residence time of the gas and of the fine particles may be on the order of 0.1 second. This would be true if the fluidized bed had no capacity to retain and distribute fine particles. If the fuel were not evenly distributed and retained by the bed so that it could react with the air passing evenly through the grid plate, the fuel would burn in the freeboard with the undesirable results shown in Figure 29.

It was shown in Section 6.5.3 that at a low bed temperature ($T < 1,850^{\circ}F$), large quantities of relatively hot carbon may appear in the freeboard. If this carbon then reacts, a temperature rise will occur in the gas space of an adiabatic combustor. In a sense, the fluidized bed itself is cooled in heating the fuel, and a large temperature difference will exist between the fluidized bed and the space above the bed.

Despite an extensive search, no usable correlation was found which would predict the fine particle residence time in a high velocity fluidized bed. The work required to produce such a correlation was beyond the scope of this study. Even with such a correlation, performance predictions would still be difficult, since the fly ash particles are continually becoming smaller as they are consumed. An additional and more practical problem is that the fine solids may be pneumatically injected at a single point, or series of discrete points. Therefore, an average superficial velocity may not adequately describe the actual gas velocity in the vicinity of the injector, and it is probably this local velocity which most influences the fine particle residence time. There is an obvious need for basic research on the behavior of fine particles in a fluidized bed composed of relatively large particles.

It is apparent that an effective Carbon-Burnup Cell design requires that the bulk of the fly ash burn within the bed and not in the freeboard. Burning within the burnup cell bed is favored by high temperatures and even fuel distribution across the cross-section of the bed. These two factors are, of course, related. If the fuel is poorly distributed, its combustion will not be complete in the bed and the bed must cool. On the other hand, if the burnup cell bed is heavily cooled--by heat exchange surface, by excess air, or by circulation of bed material from the cooler primary cell region -- combustion within the bed will also be incomplete and the bed will cool further. Carried to the extreme, poor fuel distribution and/or excessive cooling will result in the fire being extinguished.

### 7.3.2 Fly Ash Feeder Designs Tested

During the first shakedown tests with the FBM, the combustion efficiency of the CBC was extremely poor; i.e., CE ≈ 25%. During these tests, the CBC was excessively cooled by circulation of 1,500°F bed material from the FBM. An additional cause of the poor performance may have been the use of a horizontal stub fly ash injector similar to that used in the FBC (See Figure 4). We found that the temperatures near the base of the bed (3" above the grid) were lower than those 9" above the grid, (near the center of the expanded bed), and that the temperatures in the gas space directly above the bed (28" above the grid) were higher still.

A change was made to the "mushroom" feeder described in Section 5.2.2 and shown in Figure 11, and the undesirable gradient was effectively removed. Table 19 lists the temperatures in a test with the horizontal stub feeder and a test with the mushroom feeder.

TABLE 19

EFFECT OF FLY ASH FEEDER DESIGN ON
TEMPERATURE DISTRIBUTION IN A CARBON-BURNUP CELL

| Feeder Type | Temperatures, °F Horizontal Stub | Temperatures, °F Mushroom |
|---|---|---|
| Thermocouple locations (Distance above air distribution grid) | | |
| 3" (Base of bed) | 1,770 | 1,910 |
| 9" (Center of expanded bed) | 1,810 | 1,930 |
| 28" (10" above expanded bed) | 1,950 | 1,900 |

The CBC has a total cross-sectional area of 1.1 $ft^2$. Table 19 indicates that the single mushroom feeder effectively services this area. As noted earlier, it is not known whether a greater area could be served via a single mushroom. It seems reasonable to attempt to place similar feeders on about a two-foot triangular pitch to feed fly ash to a large Carbon-Burnup Cell.

## 7.4 Interchange of Bed Material

### 7.4.1 General

Under "Apparatus", Section 5, the constructional details of the FBM and CBC were described. These two sections of the integrated FBM system shared a common fluidized bed. The common perforated wall permitted bed particles to interchange between the FBM and CBC regions.

The FBM region contained the only direct-contact cooling surface, the water walls. The FBM had the sole igniter and also the only means for adding or removing bed* material from the system.

The interchange of bed material through a barrier has not received attention from investigators in fluid-bed technology, although it might be a useful tool in the design of catalytic reactors. Perforated barriers may be an important design feature of fluidized-bed boilers for the following reasons:

1) The barriers will permit a large boiler to be started with one or just a few gas/oil fired burners. A special starting zone may be provided without numerous cooling tubes; once fired, hot bed material and burning coal would flow through the barrier and ignite adjacent regions.

2) When the steam required from the boiler is about 50% of the full load capacity, the barriers make it possible to turn-off sections of the boiler without causing the bed depth in the active sections to decline.

3) The barriers will permit the use of a common bed for both primary and burnup cells, even though these two sections may operate with a 500°F difference in bed temperature.

In the FBM/CBC system, some of the proposed functions of the semi-permeable barrier were explored, within the limits of the apparatus.

It was found possible to ignite the CBC by allowing hot fluidized bed material from the FBM to interchange with the cold fluidized bed of the CBC. When the CBC bed reached about 800°F, coal could be fed and the CBC brought to operating temperature. Alternatively, the interchange could be allowed to continue until the CBC had reached about 1300°F at which point fly ash could be used as a fuel to further raise the CBC temperature. This procedure was as follows. First the FBM was ignited using a propane-fired burner while the CBC was still static. Then the CBC was fluidized. It was also found possible to shut down the CBC, continue operation of the FBM, and then restart the banked CBC.

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*A CBC bed sampling outlet pipe and valve were added later.

All of the features, and potential problems, of semi-permeable barriers should some day be rigorously explored. However, the one of major importance in the study of a Carbon-Burnup Cell is the rate of bed material interchange between the FBM and CBC. It is important to determine how this rate would affect the required temperature differential between the coal-burning and fly ash-burning regions of a fluidized-bed boiler. Using limestone as the sulfur control agent, it is likely that the primary cells will operate at 1500°F; the burnup cells might well operate at 2050°F.

### 7.4.2 Measurement of Rate of Interchange

No direct method exists for measuring the exchange rate.* Therefore an indirect method was required. One such indirect method would have involved adding a tracer material to the FBM bed, in a pulse, and then, by sampling the CBC bed, determining how rapidly the tracer concentration in the

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*Radioactive tracer techniques might be useful.

CBC reached the level in the FBM. Unfortunately the tracer experiment would have been costly to perform and was beyond the scope of this program. In observing the startup of the system, an inexpensive tracer-like experiment suggested itself. By igniting the FBM and bringing it to a steady-state temperature while the CBC was inactive, a source of hot bed material would be available on only one side of the barrier. Then, by fluidizing the CBC and determining how rapidly its bed temperature increased, a measure of the interchange rate could be obtained.

The derivation of equations describing the transient heating and the details of the experiments are given in Appendix J. In two experiments, in which the area of the opening between regions was different, values of the interchange rate were obtained.

A value of ≈ 12,000 pounds per hour of bed material interchange per square foot of open area was found. A value of 10,000 pounds per hour per $ft^2$ should be used in estimating the time to raise the temperature of a cool bed to near that of an adjacent hot bed.

### 7.4.3 Effect of Interchange of Bed Material on Carbon-Burnup Cell

As noted earlier, Section 5.2.2, the initial open area in the wall separating the FBM from the CBC was 90 square inches. With this open area, it was not possible to achieve a CBC bed temperature more than about $200^{\circ}$F above that in the FBM. By reducing this open area, finally to 2 square inches, it was possible to achieve as large a differential as desired. For example, in FBM test B-10 a differential of $450^{\circ}$F was achieved.

As the tests in the FBM proceeded, the carbon steel plate, used to reduce the open area, deteriorated and by Test B-13 a 300°F differential was the maximum possible. The plate was changed to ASTM 446 and the differential restored; for example, in Test B-13 the FBM operated at 1590°F and the CBC at 2000°F.

Since an open area of 2 $\text{in}^2$ between the FBM and CBC was found acceptable, it is assumed that for a larger Carbon-Burnup Cell, rapid ignition and equal bed depths may be achieved with a suitably larger opening.

The most straight forward scaling factor is to keep the ratio open area: plan area constant. For the CBC this was

$$\frac{2\ \text{in}^2}{1.1\ \text{ft}^2} = 1.82\ \text{in}^2/\text{ft}^2$$

Thus, for the Carbon-Burnup Cell, described in Section 6.5.4, with a plan area of 35.4 $\text{ft}^2$, an open area in the tube wall between primary cell and burnup cell would be 64.5 $\text{in}^2$ (0.45 $\text{ft}^2$). This open area would then result in the interchange of 5400 lb/hr of bed material between the primary and burnup cells. For a primary cell temperature of 1500°F and burnup cell bed temperature of 2050°F this interchange would remove 650 K-Btu/hr from the burnup cell which is negligible when compared to the total energy loss of 32,800 K-Btu/hr.

## 7.5 Prevention of Carryover of Jetted Particles from a High Velocity Fluidized Bed

The addition of the CBC to the existing FBM required compromises in the location and design. The decision was made to locate the CBC at the rear of the unit so that access to the larger FBM would not be more difficult and to avoid cutting pressure parts. This location limited the effective height of the CBC to 48" (the height above the air distributor at which the exhaust duct was placed, See Figure 9).

Although circumstances forced a small freeboard in this instance, it is desirable that a Multicell Fluidized-Bed Boiler have an open freeboard that is far smaller than the open freeboards most often used in fluid-bed roasters, etc. The test work in the CBC therefore afforded an opportunity to explore the potential for small freeboard.

While the low height of the CBC did not interfere with the combustion process within the bed itself, it did cause the loss of bed material. The bed material would be carried over and would appear in the dust collector discharge. Since the CBC bed was in communication with the FBM bed, the bed depth in the smaller CBC would remain relatively constant despite the depletion of bed material. The cause of this carryover has been discussed in detail in Appendix L.

Prevention of carryover in fluidized beds is normally accomplished by insuring that the exhaust is above the point in the vessel where a relatively high density of jetted particles exists. In the case of the CBC, this was not possible without a major change to the FBM's steam drum. (This change was subsequently made and will be discussed in a future report).

An alternative to a large, open freeboard is to install some form of baffle at the top of the vessel. If the particles strike this baffle and thus lose their vertical momentum, those particles large enough to return to the bed against the average gas velocity will return. The use of a louvered baffle in this application was described by Loshe and Commander (25) in reporting on the fluidized-bed calciner for radioactive wastes built at the National Reactor Testing Station. The installation of such a louvered baffle in the CBC was being contemplated when a visit to BCURA Industrial Laboratories, under OAP sponsorship, revealed BCURA's application of an array of horizontal heat-exchange tubes above the bed of a fluidized-bed combustor which accomplished the same knockout function. Such an array was installed in the CBC.

By providing a surface on which large particles may lose their kinetic energy, the baffle screen allows these particles to fall back into the bed. The baffle screen does not prevent elutriation of fine material, that is particles whose density and diameter give them a terminal velocity below the actual gas velocity. However, if the baffle screen is made up of heat exchange surfaces, the gas temperature will drop and the gas velocity, which is proportional to gas temperature and inversely proportional to free area, will also drop, permitting some of the marginal particles to return to the bed.

The first knockout array, studied in the current program, consisted of seventeen one-inch O.D. water-cooled tubes arranged in four rows on a triangular pitch. The arrangement was "optically dense"; i.e., such that, when viewed from above the baffle screen, the bed could not be seen. Therefore there was no line-of-sight path for a jetted particle to travel into the exhaust duct without striking a tube. The tube arrangement is shown in Figure 9.

After the installation of this baffle screen, large bed particles (-8+22 mesh) no longer appeared in the cyclone discharge. The size distribution was then similar to that in Figure 26.

The cooled screen was such an effective heat exchanger that the CBC bed had excessive cooling and the combustion efficiency suffered. The heat exchanger performance of the cooled screen is discussed below.

A second baffle screen was also tested. This baffle was identical in arrangement to the first, but was made up of uncooled ASTM 446 rods. This screen also prescribed the carryover of ash bed particles. The uncooled screen prevented chilling the bed.

The designs tested and found serviceable consisted of four rows of staggered tubes. Additional rows, more widely spaced, may be used to avoid erosion by the heavy dust loading in the Carbon-Burnup Cell exhaust. A maximum velocity of 30 feet per second is recommended for this application based on conservative, conventional boiler practice. For a high-velocity Carbon-Burnup Cell, operating at a superficial gas velocity of 15 fps, a free area reduction of 50% due to the baffle screen would yield the desired result.

We recommend that an open freeboard space of no less than 36 inches be provided between the top of the expanded bed and the entrance to the convection bank or cooled baffle screen. In the CBC, the comparable distance was only 10 inches. The 36" distance, provided in the FBM, is considered the minimum for completion of combustion occurring in the freeboard.

We recommend that any fluidized-bed boiler make use of the baffle-screen concept so that the freeboard height will not be as great as that required for disentrainment of bed particles by gravity alone.

## 7.6 Performance of a Baffle Screen As a Heat Exchanger

As described in the previous section, an array of water-cooled tubes was installed in the CBC to reduce carryover of bed material. The screen, consisting of 1" tubes arranged on a triangular pitch, was located between the 32" level (i.e., 32" above the air distribution grid) and the 39" level. (See Figure 9). The expanded bed depth was about 20" for most tests. Therefore, there was, on the average, a 10-12" gap between the dense phase fluidized bed and the bottom of the baffle screen. As noted earlier, there is considerable splashing in the dense-phase bed so that hot particles would hit the tubes.

In designing the screen, it was anticipated that the effective average heat transfer coefficient would be no higher than about 18 $Btu/ft^2hr^oF$, calculated for convection from the gas and radiation from the bed to the viewed tubes. If the coefficient were 18, a heat absorption rate of ≈ 150,000 Btu/hr would be anticipated for the tube array.

When actual absorption was determined by measuring the coolant flow and temperature rise, a value of over 300,000 Btu/hr was indicated. Overall coefficients of ≈ 35 $Btu/ft^2hr^oF$ were calculated, using the bed temperature as the gas side temperature in the basic heat transfer equation. If actual gas temperatures entering and leaving the tube bundle were used, the coefficients compare to ≈ 50 $Btu/hr\ ft^{2o}F$ to a single tube immersed in the fluidized bed.

This result demonstrates that a heat exchanger placed above the dense phase of a high-velocity fluidized bed will experience a heat flux almost as great as if it had been immersed in the bed. This is significant for the design of a Multicell Fluidized-Bed Boiler, since a much lower fan power penalty is paid for surface located above the bed than in raising the bed level so as to immerse more surface. While the height of this baffle screen above the bed could not be varied, it was found in more recent tests that the heat transfer coefficient decreased as the distance above the bed increased.

In the absence of tests with a movable heat transfer bundle,* a film coefficient ≈ 35 $\mathrm{Btu/ft^2hr^oF}$ should be used for heat exchangers located as in the CBC ( ≈ 12" above the expanded bed). This coefficient should apply only to about the first four rows in the exchanger. Beyond the fourth row, the fluidized-bed should no longer affect the heat exchange coefficient and normal convective transfer coefficients may be calculated and used in design.

## 7.7 Performance of Integrated FBM/CBC System

The performance of the FBM and the appended CBC which was developmental in nature was poor during most of the test program described here. Test data, listed in Appendix K, are summarized below in Table 20. It may be seen that in only two tests did the system operate at an integrated combustion efficiency of about 99%, and this was when both the FBM and CBC burned carbon efficiently. Burnup calculations are uncertain when the ash output is not weighed.

Because of a number of problems, some of which were outlined in the preceding sections, it was not found possible

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*Tests with single heat-transfer probes are not considered adequate to provide engineering data for design of heat exchangers made up of an array of tubes. One reason is that aerodynamic similarity is inadequate.

## TABLE 20

## PERFORMANCE OF FBM, CBC AND SYSTEM IN CURRENT TEST SERIES

| | COMBUSTION EFFICIENCY, % | | |
|---|---|---|---|
| TEST NO. | IN FBM[1] | IN CBC[1] | IN SYSTEM[2] |
| B-5 | -- | 82 | -- |
| B-7a | -- | 75 | -- |
| B-7b | -- | 78 | -- |
| B-8 | 76 | 75 | 94 |
| B-9 | 76 | 74 | 94 |
| B-10 | 78.4 | 72 | 94 |
| B-11 | 75 | 53 | 89 |
| B-11b | 61 | 73 | 90 |
| B-12 | 78.5 | 67 | 93 |
| B-12b | 67 | 57 | 87 |
| B-12e | 73 | 35 | 84 |
| B-13c | 83.2 | 60 | 93.3 |
| B-14 | 81.5 | 95 | 99.1 |
| B-15 | 80.5 | 93 | 98.6 |
| B-16b | 79 | -- | -- |
| B-17 | 83 | -- | -- |

[1] CE = (Carbon Burned/Carbon In) x 100

[2] CE = (Heat Released in FBM and CBC/HHV of Coal Feed) x 100

to perform the numerous parametric studies with the FBM which might have provided additional useful data. The desired statistically designed test program was not conducted with the FBM.

The operational problems found in the FBM during this test series were as follows:

1) Debugging of the vertical coal feeder (Described in Section 7.2).

2) The new coal handling system, which included a ten-foot length of inclined screw, produced excessive fines in the coal feed. The coal delivered at the outlet of the feeder was captured in a cold test and found to contain 60 wt. % of -16 mesh material. This is roughly twice the fines content of the coal used in past work (1, 4). In operation, the fines would be carried out of the FBM without completely burning; thus, the inability to reach the 85% carbon combustion efficiency expected from the FBM. The coal handling system could not be repaired or replaced within the time and funds available.

3) There was evidence that the grid plate had warped in the vicinity of the CBC and that some air was bypassing the bed. This was indicated by signs of erosion. The symptom of bypassing of this type, would be a lower carbon combustion efficiency, higher hydrocarbons in the flue gas and poorer sulfur removal by limestone at coal/air ratios that in the past had produced better results. The plate was caulked but this was only a temporary repair. Recently (1971) a metal seal was installed that may have reduced the leakage.

The operational problems found in the CBC during this test series were as follows:

1) The excessive cooling due to the high rate of interchange of cool bed material from the FBM (See Section 7.4).

2) The poor performance of the original fly ash

injector (See Section 7.3).

3) The excessive carryover of bed materials caused by the inadequate freeboard (See Section 7.5).

4) The excessive cooling due to the unexpected high heat flux into the water-cooled baffle screen (See Section 7.6).

As noted above, the last two tests in the integrated system were highly efficiently, and when most of the operational problems had been eased both the FBM and CBC performed acceptably well.

The effectiveness of the performance model, derived in the FBC, was tested with the results of CBC tests. The observed and calculated combustion efficiencies are shown in Figure 38. It is seen that the results of only two tests were predicted precisely. It is unfortunate that time was not available to perform a series of tests with the properly operating apparatus over the entire efficiency range (75% to ~90%) to fully test the performance model.

## 7.8 Sulfur Capture with Fine Limestone Injection

Reference 1 describes the procedures and results of a number of tests in which fine limestone was added to the FBM before the CBC was appended. The details of these tests will not be repeated here. The addition of the CBC to the FBM required major changes to the apparatus. Among these changes was a new coal injector; another was a new limestone injector. These have been described in Section 5. The procedures used in the current test series were identical to those described in (1). Presumably then, only the equipment changes could have affected the sulfur removal performance of the FBM.

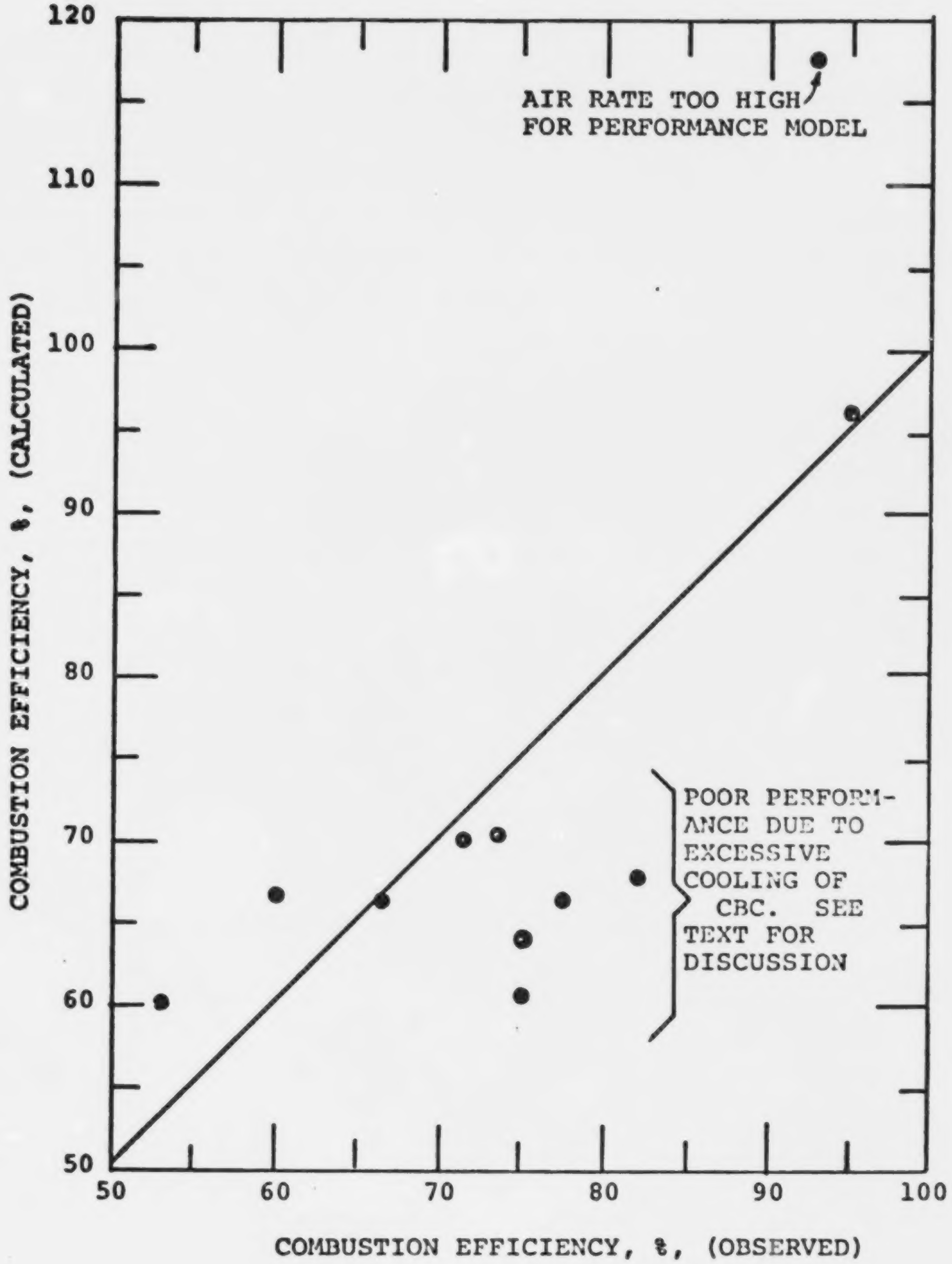

**FIGURE 38** COMPARISON OF OBSERVED AND CALCULATED COMBUSTION EFFICIENCY OF CBC

When fine sorbent was injected into the FBM, it would pass through the unit to the FBM dust collector. Then, along with the carbon-bearing fly ash, the partially-sulfated limestone would be injected into the CBC, pass through the CBC, and be collected by the CBC dust collector. From the CBC dust collector hopper, this mixture of "burned out" fly ash and lime by-products could be wasted or reinjected into the boiler.

Figure 39 compares the data previously obtained with some of the data obtained in this program. It may be seen that more limestone was required in the current test series to achieve the same level of $SO_2$ reduction. No reason for these poor results was apparent, and no attempt was made to alter the system so that the formerly high efficiency could be restored. It is believed that the currently installed additive injectors are less effective than those used in (1) and that much of the additive leaves the bed in one location without reacting with sulfur. Poor coal distribution might also account for poor sulfur removal.

Emphasis is now being given to a regenerative limestone process (Section 4.3.3) instead of once-through fine limestone injection. Therefore, no additional effort in this area could be justified.

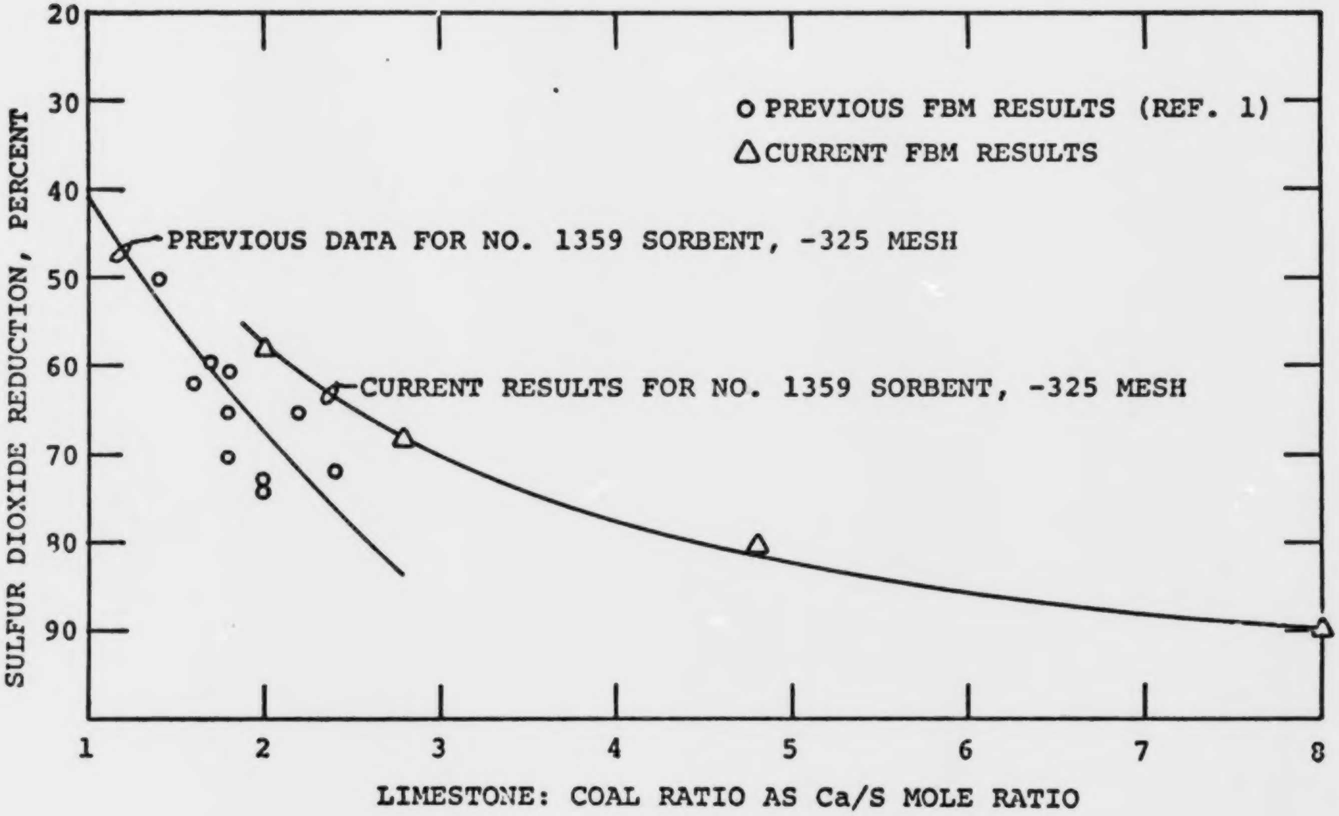

FIGURE 39 SULFUR DIOXIDE REDUCTION WITH FINE LIMESTONE INJECTION INTO FBM

# 8. ECONOMICS OF THE CARBON-BURNUP CELL

## 8.1 Perspective

The detailed design and pricing of a Multicell Fluidized-Bed Boiler were beyond the scope of the current program (See Reference 4 for detailed cost data on an early Multicell Fluidized-Bed Boiler design). A cost estimate is not provided.

In the current study, devoted almost entirely to the Carbon-Burnup Cell portion of the Multicell Fluidized-Bed Boiler, the most significant question of economics concerns optimizing the Carbon-Burnup Cell as a part of the boiler. Unfortunately, as noted in Section 6.4.8, the performance model is not sufficiently sophisticated to permit a meaningful evaluation of the comparative capital or operating costs of various alternative Carbon-Burnup Cell designs. More specifically, for example, it is not meaningful to compare the costs for operation with a bed depth of 24" and a bed depth 48" since Equation 1 would give an incorrect result for the 48" bed depth. For this reason it is not possible, using the information available, to optimize the Carbon-Burnup Cell alone.

Despite the performance model's deficiency, it is not recommended that any additional effort be devoted to extending its range to permit such optimization. The existing model describes with sufficient precision how a Carbon-Burnup Cell should be designed.

The material which follows will show how the Carbon-Burnup Cell performance model plus data on the combustion efficiency in the primary cells of the multi-cell boiler may be used to optimize the design of the entire unit. The analysis is based on the premise that the unit is optimized when fuel costs are minimized and that fuel costs are minimized when the thermal efficiency of the boiler is maximized. Implicit in this statement are assumptions that:

1) The capital cost of burnup cell and primary cell, expressed in dollars per Btu of heat release rate, are equal. Since the primary cells may be operated at heat release rates of from 800,000 to 1,200,000 Btu/hr $ft^2$ of grate area, while the burnup cells would be operated at 500,000 to 800,000 Btu/hr $ft^2$, this assumption is only approximately true.

2) The auxiliary power requirements (mainly for the forced draft fans) expressed as KWH per Btu of heat release are equal for the primary and burnup cells. This assumption is also not strictly true since from 10 to 20% more air, per Btu of heat release, will be used in the burnup cells than in the primary cells.

The specific capital and auxiliary power costs of the burnup cells are therefore somewhat higher than the same costs in the primary cells. However, it is felt that the $1^o$ cell - burnup cell system can be optimized without regard to these differences in cost because:

1) The capital costs which are heat-release-rate and excess-air-rate dependent are in the air distributor and in fans, ducts, breeching and dust removal equipment. Heat transfer surface is not included in this category. A 50% change in heat release may result in only about a 10% change in the capital cost, $/Btu of heat release, of these items.

2) The auxiliary power required for fans by the Multicell Fluidized-Bed Boiler will be on the order of 1.0

to 1.5 KWH per $10^6$ Btu of heat released. For a 40% efficient power station, 1 KWH - 8550 Btu of fuel input or about 0.85% of $10^6$ Btu. The difference in specific auxiliary power requirements of burnup cell and primary cell are therefore negligible.

3) It will be shown that an optimum design is achieved when the primary cells burn about 90% of the fuel and the Carbon-Burnup Cells the remaining 10%. The curves, to be developed, are flat and no sharply defined optimum is established.

On the basis of these three arguments it is felt that the simplifications do not detract from the essential validity of the premise that the system is optimized when the boiler's thermal efficiency is maximized.

## 8.2 Analysis

In numerous tests in fluidized-bed coal combustors, it has been found that a sizable fraction of the fuel value of the input coal may be lost as unburned carbon in fly ash. This loss may range from 5% to 20% of the input carbon, or about 5% to 15% of the input energy value of the fuel when the hydrogen content of the coal is accounted for.

Two basic methods of recovering the fuel value in the fly ash have been proposed. In one design concept, the boiler is operated at low air rates, (≈ 2 to 4 fps superficial gas velocity) and the collected fines which contain unburned carbon are recycled back to the fluidized bed from which the fines originated. In a second design concept, which is favored here, the fluidized-bed boiler operates at high air rates (≈ 10-15 fps). Here the collected, carbon-bearing fines are recycled back to a special region of the fluidized bed in which a relatively high residual oxygen and temperature ( > 1950°F) are maintained by providing less heat-transfer surface per unit fuel burned than in the coal-burning regions.

The data gathered in an earlier study (4) established that the carbon in fly ash could not be effectively burned in a fluidized bed at 1500-1700°F with 3% residual oxygen in the flue gas at high (≈ 6 fps) air rates. Pending the disclosure of an invention which provides an alternative to the Carbon-Burnup Cell, such a region must be provided if high levels of combustion efficiency are to be achieved in an economically viable fluidized-bed boiler. As examples of boilers which we do not consider economically viable, consider those with high excess-air levels and those with unrealistically low gas velocity.

Figure 40 shows the response, carbon combustion efficiency, for the carbon in coal, as a function of the major pseudo-control variable, residual oxygen in the flue gas. Although residual $O_2$ is itself a response to fuel rate, air rate, fuel reactivity, etc., in actual operation, residual $O_2$ is treated as a control. Figure 40 shows how combustion efficiency increases with increasing residual $O_2$.

Note that these data, obtained in different apparatus over the past several years, represent coal burning tests, not tests in which fly ash was the fuel. The combustors were operating in the manner of a primary cell, not as a carbon burnup cell. The fuel was 1/4" x 0, not -1 mm.

A power curve was fit to the data which appears intuitively correct although its mathematical form is not the most appropriate. This curve has the form

$$CE = a\,(RO)^{b}$$

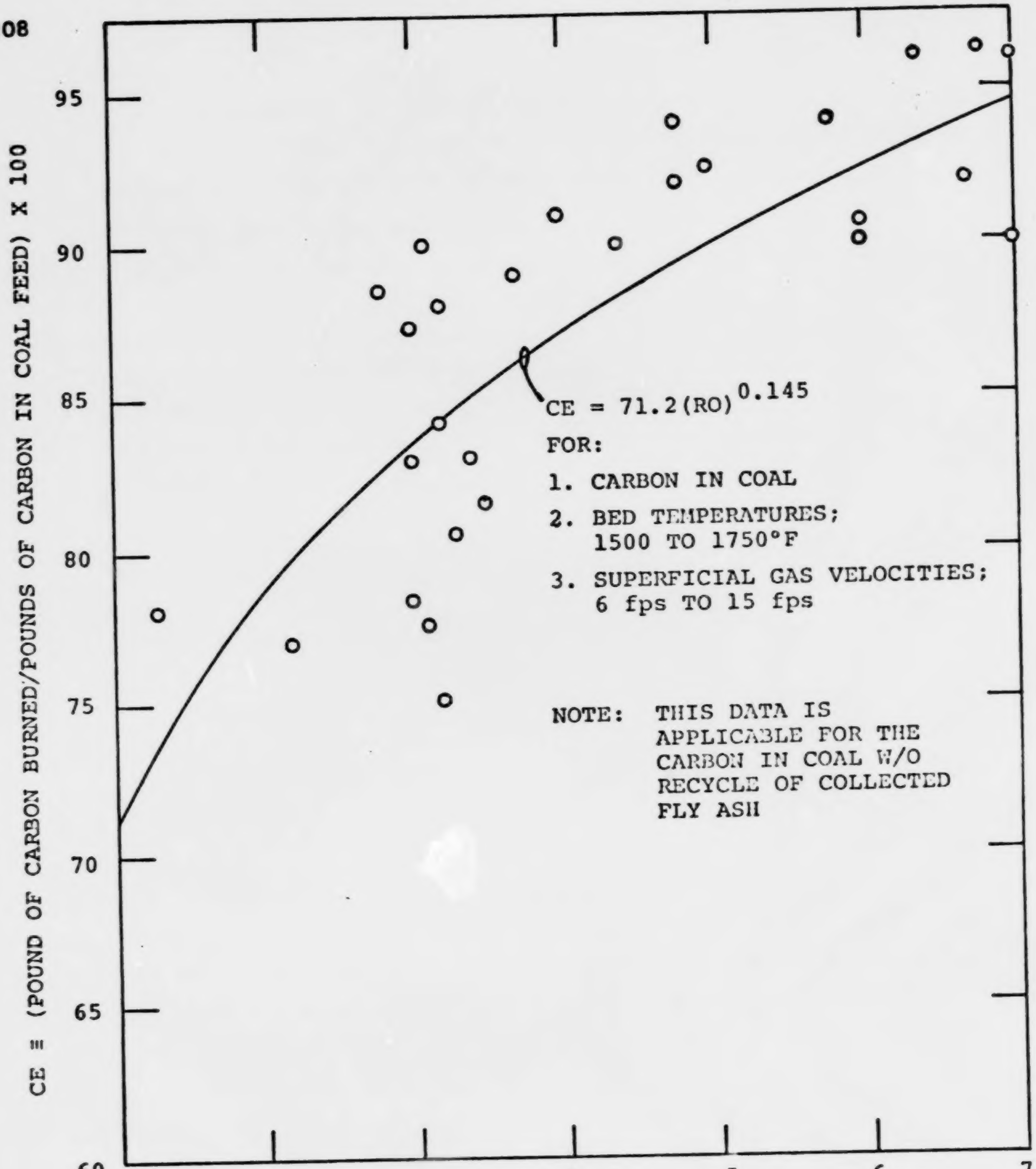

FIGURE 40 CARBON COMBUSTION EFFICIENCY IN A COAL-FIRED FLUIDIZED-BED COMBUSTOR

where

$$CE = \left(\frac{\text{weight of carbon burned}}{\text{weight of carbon in coal fed}}\right) \times 100 \qquad (41)$$

$RO$ = residual $O_2$, vol. %

with coefficients

$a = 71.2$ and
$b = 0.145$

for RO between 1.3 and 7.0, with standard deviation $s = 4.1\%$.

The relationship

$$\frac{100}{100-CE} = 2.6 + e^{(0.39 \times RO)} \qquad (42)$$

could be used for extrapolation to high values of RO. For RO between 1.0 and 8.0, Equations (41) and (42) give identical results.

With this expression and other data obtained in this study, it is possible to determine:

1. The optimum operating point for a fluidized-bed boiler containing primary cells only and without fly ash recycle of any kind.

2. The optimum division of duty between the coal-burning primary cells and fly ash-burning Carbon-Burnup Cells of a Multicell Fluidized-Bed Boiler for minimum fuel costs.

### 8.2.1 Evaluation of a Fluidized-Bed Boiler Without Fly Ash Cycle

Example 1.

Problem: Using Equation (41), determine the optimum level of residual oxygen for a fluidized-bed boiler consisting only of primary cells and in which the fly ash is not recycled.

Solution: Determine the residual oxygen level at which the sum of the energy loss in unburned carbon plus that in the flue gas is minimized.

Data:

a. The coal described in Table 16 (Page 150) will be burned having an as-fired heating value of 12,500 Btu/lb.

b. Flue gas will be discharged at 400°F with an enthalpy of 81 Btu/lb of gas.

c. Carbon in fly ash has a fuel value of 14,100 Btu/lb.

d. The latent heat loss, 49,000 Btu/100 lbs of coal, is a constant and need not be considered in finding the optimum.

Results:

Figure 41 shows (1) the heat loss due to unburned carbon as a function of residual oxygen (RO) level calculated from Equation 41; (2) the sensible heat loss in 400°F flue gas as function of RO; and the total heat loss being the sum of (1) and (2).

The results plotted in Figure 41 reveal that the total loss does not reach a minimum even at RO = 8%. (The computations showed a minimum at about RO = 10% but the loss is not much different than at 8%). The total loss for a typical highly efficient pulverized coal (p.c.) furnace, also with 400°F exhaust temperature, is also shown on the figure. The range shown is for 15 to 20% excess air and a 0 to 1.0% combustible loss typical of p.c. units.

Note that the minimum loss for the fluidized-bed boiler is higher than that for the pulverized coal boiler.

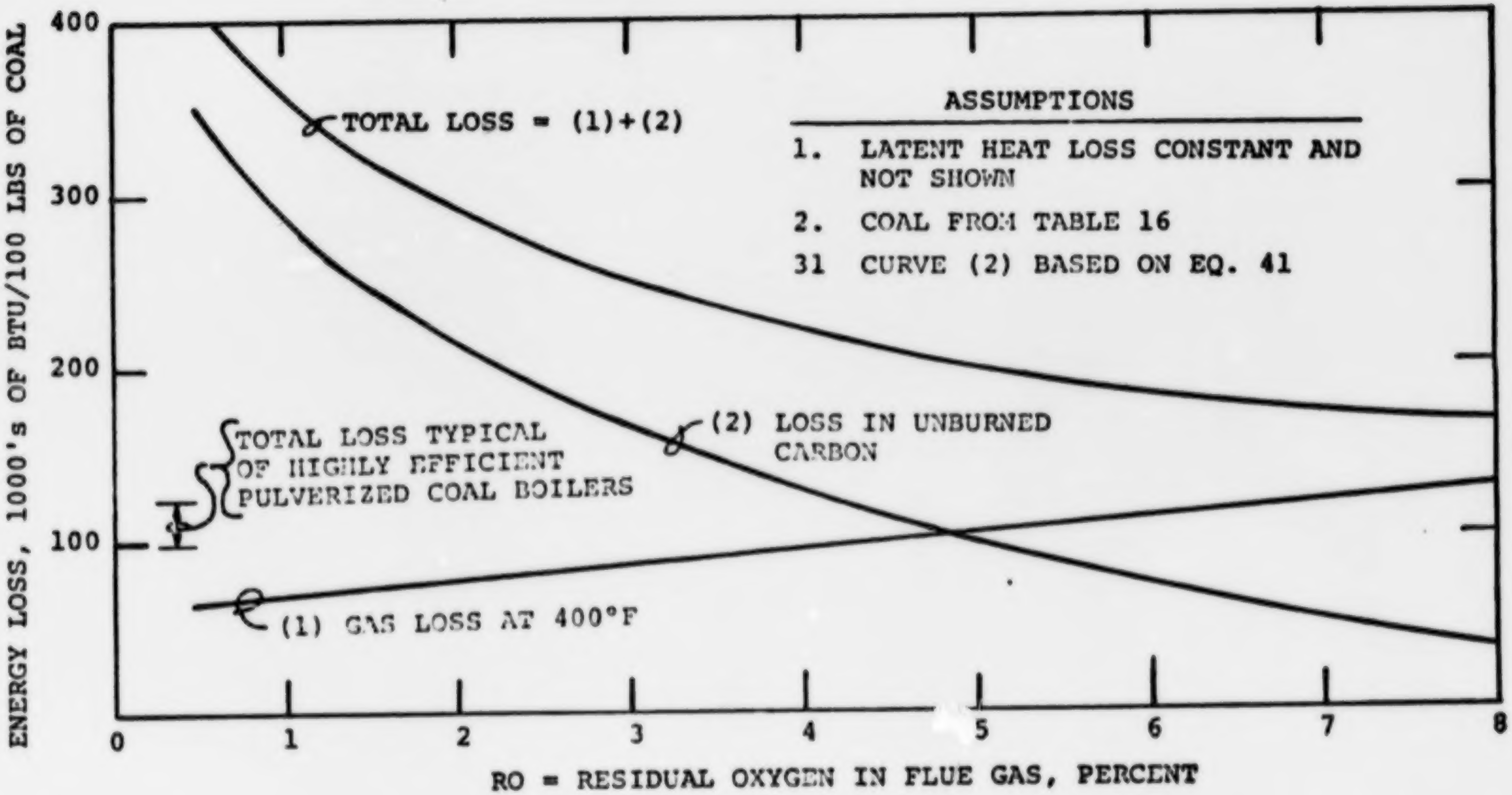

FIGURE 41 THERMAL LOSSES FROM FLUIDIZED-BED BOILER OPERATING WITHOUT RECYCLE OF CARBON-BEARING FLY ASH

Including the latent heat loss, the fluidized-bed boiler achieves a maximum thermal efficiency of 83.4% while the p.c. unit is 86.5% efficient.

A Carbon-Burnup Cell is therefore required if the fluidized-bed boiler is to operate with a thermal efficiency comparable to that of the p.c. unit.

With no fly ash recycle, the minimum loss point moves toward even greater values of residual oxygen as the exhaust temperature is reduced.

### 8.2.2 Evaluation of a Multicell Fluidized-Bed Boiler: Optimum Duty Split Between Primary Cells and Carbon-Burnup Cells

Example 2.

Problem: In the case of a Multicell Fluidized-Bed Boiler in which the carbon lost by the primary cells is burned in the Carbon-Burnup Cells, determine if an optimum exists in the division of duty between the coal-burning cells and the fly ash burning cells.

Solution: Determine the residual oxygen level in the coal burning regions at which the total energy losses are minimized.

Data:

a. The coal described in Table 16 will be burned.

b. Flue gas will be discharged at 400°F with an enthalpy of 81 Btu/lb from the coal-burning regions (5% moisture) and 79 Btu/lb ($<$ 1% moisture) from the fly ash burning regions.

c. Carbon in fly ash has a fuel value of 14,100 Btu/lb.

d. The latent heat loss, 49,000 Btu/100 pounds of coal, is a constant and need not be considered in finding an optimum.

e. The Carbon-Burnup Cell has the following parameters:

(1) Fuel input rate = C'(14,000) = 500,000 $Btu/ft^2hr$

(2) Bed temperature, T = $2000^{\circ}F$

(3) Bed depth, H = 24"

(4) Combustion efficiency, CE = 85, 90, and 95%

Procedure:

1. Assume a value of RO from the primary cells, calculate CE for carbon in coal using Equation 41. From this compute the composition of the primary cell fly ash, % C, % inert to be fed to the burnup cells.

2. Compute the weight of air required by the primary cell at the assumed value for RO. For these calculations assume complete combustion of the hydrogen and sulfur content of the coal. The air for carbon and residual oxygen is computed from Equation 21. Compute the total weight of flue gas produced and the sensible heat loss at $400^{\circ}F$.

3. Using Equation 1 and for the conditions given above in the data for Carbon-Burnup Cells, compute the air-to-carbon ratio A'/C' to produce the specified burnup cell combustion efficiency. Compute the flue gas weight and sensible heat loss at $400^{\circ}F$. Compute the heat value for unburned carbon.

4. For one value of RO in the primary cell flue

gas and one value of burnup cell efficiency compute the total heat loss (neglecting the latent heat loss) which is:

(a) the sensible heat loss of primary cell flue gas; plus,

(b) the sensible heat loss of burnup cell flue gas; plus,

(c) the heating value of unburned carbon.

5. Repeat computations for a range of values of primary cell RO and burnup cell CE's.

Results: The sum of thermal losses, dependent on the residual oxygen level, from the two regions of a Multicell Fluidized-Bed Boiler are shown graphically in Figure 42 (add 49,000 Btu to the values shown to include the latent heat loss). It may be seen that an optimum does exist at about 2-1/2 to 3-1/2% residual oxygen in the primary cell flue gas. This range in residual oxygen may be related to the duty on the Carbon-Burnup Cell. At RO = 2-1/2% in the primary cell flue gas, the Carbon-Burnup Cell fires 14.5% of the total fuel input; at RO = 3-1/2%, the Carbon-Burnup Cell has 11% of the total duty. Total fuel input is 1,255,000 Btu/100 pounds of coal. The total loss shown for the p.c. unit of typical efficiency is the same as in the preceding example.

A second case (more heat recovery equipment used so the flue gas exhausts at 250°F) is shown in Figure 43. A slight shift in duty to the coal-burning regions is indicated here. In the case of a 250°F gas discharge temperature, the Carbon-Burnup Cell should burn 9 to 12% of the fuel. However, the curves are so flat that little penalty in efficiency is paid in optimizing on criteria other than duty division.

Thermal loss for a p.c. unit is not shown for a 250°F exhaust since low temperature corrosion, due to $H_2SO_4$,

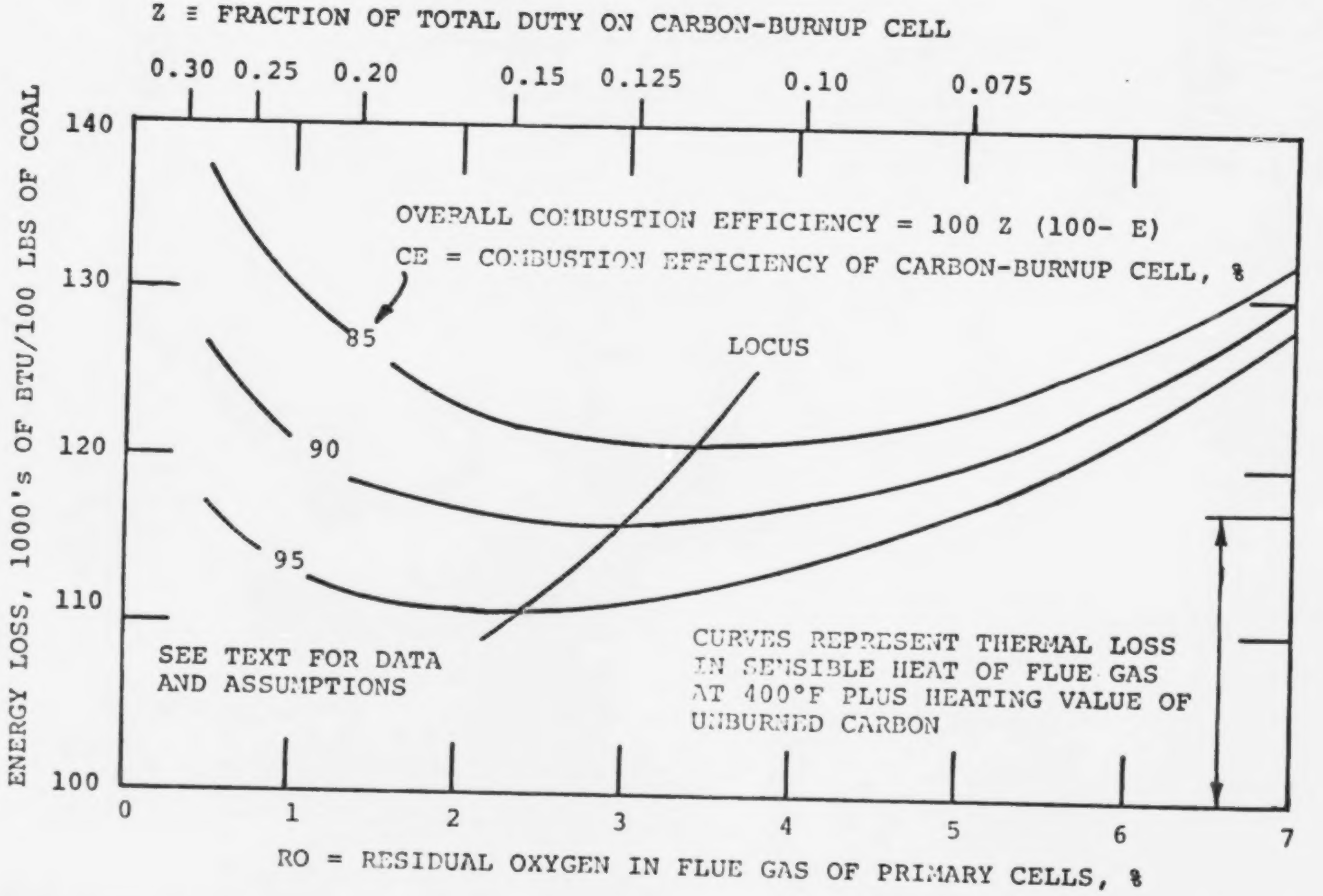

FIGURE 42 THERMAL LOSSES FROM MULTICELL FLUIDIZED-BED BOILER - EXHAUST TEMPERATURE 400°F

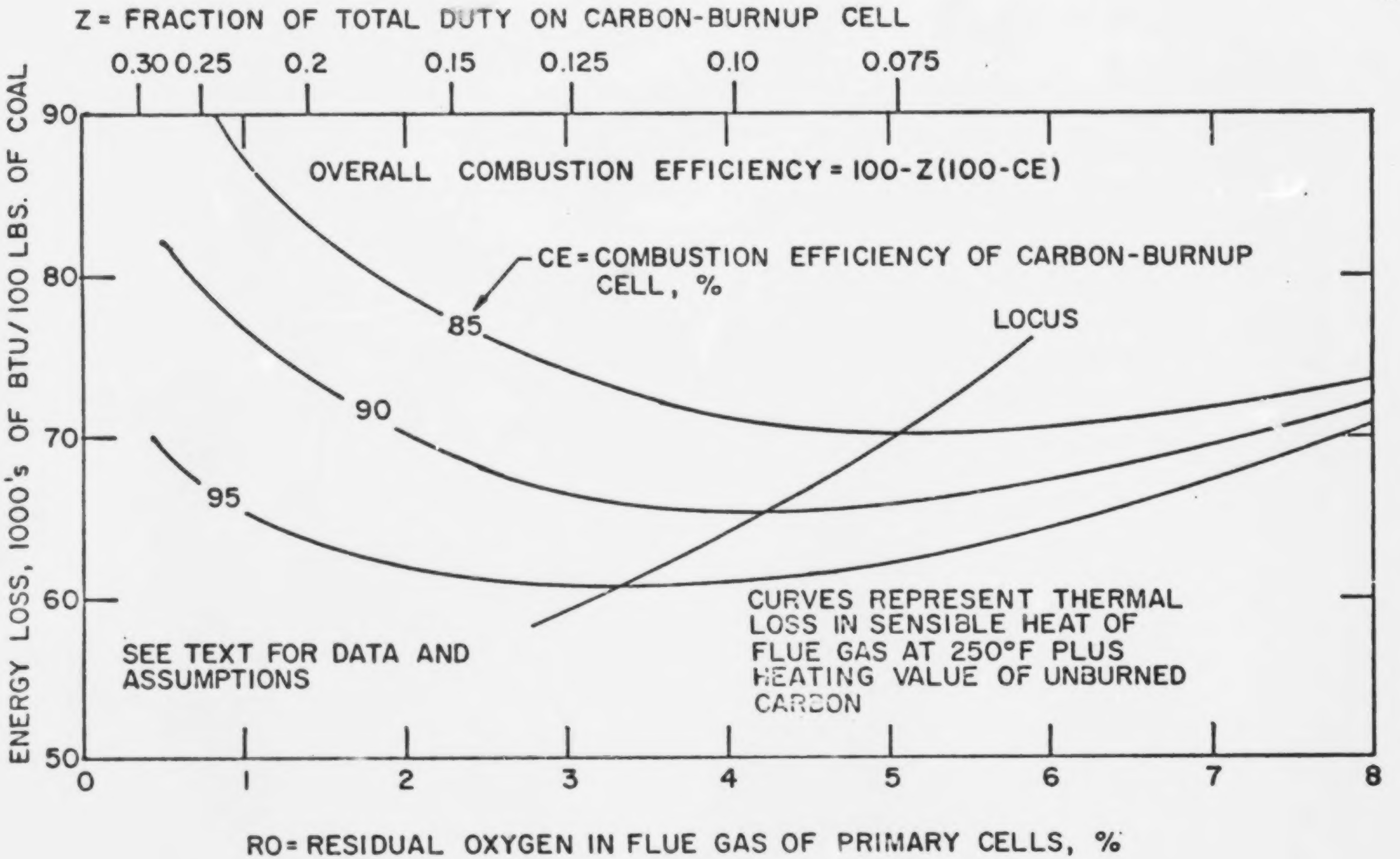

FIGURE 43 THERMAL LOSSES FROM MULTICELL FLUIDIZED-BED BOILER-EXHAUST TEMPERATURE 250°F

is avoided by operating at an exhaust temperature in excess of 300°F.

Conclusions: It is seen that with a Carbon-Burnup Cell, fuel consumption in a Multicell Fluidized-Bed Boiler will be comparable to that of a conventional pulverized-fuel boiler at a gas exit temperature of 400°F. Since lower grade, and hence lower cost, fuels may be burned in the fluidized-bed boiler, fuel costs will be lower than in the pulverized-fuel boiler.

If the gas discharge temperature is relatively low, $220 \geq T_{gas} \geq 300^{\circ}F$, the optimum duty division between the primary cells and the burnup cells covers a wide range and the selection should be based on other criteria. For hot discharge, $\geq 350^{\circ}F$, a slightly more pronounced optimum occurs between 10-15% duty on the Burnup Cell. If fuel costs are relatively high and sulfur control effective, it may be economically feasible to build Multicell Fluidized-Bed Boilers with an exhaust temperature of 250°F or even lower. The overall thermal efficiency at 250°F exhaust would be 90.5 to 91.5% depending on the efficiency of the Carbon-Burnup Cell for the coal shown.

## 9. SUMMARY OF DESIGN CRITERIA FOR CARBON-BURNUP CELL

The bulk of this report has been devoted to a discussion of the design of a Carbon-Burnup Cell. The purpose of this section is to summarize the conclusions on optimum design.

Combustion efficiency and minimum particulate emissions are achieved by operating a high bed temperature. With reliable controllers, consistent operation at 2050°F appears feasible. With the development of a technique for continuously removing larger agglomerates, operation at bed temperature exceeding 2050°F would appear desirable.

The model equations of Section 6 may be used to optimize the Carbon-Burnup Cell for a particular application.

Removal of energy by contacted and viewed heat exchange surface should not exceed about 25% of the energy release.

Air rates consistent with high superficial velocities (10 to 15 fps) may be used and are desirable for a compact design.

The unit should be designed to fire more fly ash than is anticipated so that under no reasonable circumstances will carbon-rich fly ash accumulate in the hoppers. The energy deficiency which follows from this design arrangement is to be made up with coal. No need exists to operate with the Carbon-Burnup Cell as a separate system. It is an integral part of a Multicell Fluidized-Bed Boiler and should share a common bed with the primary cells. If fine lime were injected to reduce sulfur oxide emissions, the Carbon-Burnup Cell should operate with a residual oxygen level ≥ 3% to minimize decomposition of partially sulfated limestone.

A static bed depth of over 10" is required for reasonably efficient operation; a bed depth of 18" to 24" is an appropriate design range.

## 10. REFERENCES

(1) Characterization and Control of Gaseous Emissions from Coal-Fired Fluidized-Bed Boilers, Interim Report, October 1970, for Division of Process Control Engineering, National Air Pollution Control Administration, Department of Health, Education and Welfare, by Pope, Evans and Robbins, PB198 413.

Report covers mostly fine limestone injection into fluidized bed combustors.

(2) U. S. Patent No. 3,508,506, Process and Apparatus for Reduction of Unburned Combustible in Fly Ash.

The term, "Carbon-Burnup Cell" was coined to identify the methods and apparatus described in this patent. Since the words "Carbon-Burnup Cell" refer to a specific design concept they should always be capitalized.

(3) Serial No. 66,724 filed U. S. Patent Office August 25, 1970.

The "$SO_2$ Acceptor Process" described in this patent application is a method of controlling $SO_2$ emissions from any boiler but most likely fluidized-bed boiler in which a gas containing more than about 5% $SO_2$ is produced for use or neutralization. The patentable features pertain to the methods of making this gas at relatively low cost and free of potential pollutants (other than $SO_2$).

(4) Development of Coal-Fired Fluidized-Bed Boilers prepared for Office of Coal Research, Department of the Interior, by Pope, Evans and Robbins, Interim Report, Office of Coal Research, Research and Development Report No. 36, Contract No. 14-01-0001-478. Available from Superintendent of Documents.

Report covers experimental and design studies for compact fluidized-bed boilers carried on by Pope, Evans and Robbins from 1965 through 1969. Describes apparatus used in work reported here.

(5) U. S. Patent No. 1,687,118 (October 9, 1923 and D.R.P. 437,970 (1922).

Winkler describes the invention of fluidized bed combustion and gasification.

(6) U. S. Patent No. 1,984,340 (December 18, 1934, reissued August 6, 1940).

Odell describes the invention of fluidized-bed catalytic reactors. Odell stated recently (1970) in a personal communication that he had considered the feasibility of a coal-fired fluidized-bed boiler in the 1930's but had thought that coal ash clinkers would form and make such a system unworkable.

(7) U. S. Patent No. 2,509,866 (May 30, 1950).

Hemminger describes how a coal-fired fluidized-bed boiler would operate. He recognizes that it would not be possible to burn all of the coal in a single stage.

(8) Godel, A. A., "Ten Years of Application of Fluidized-Bed Combustion of Coal" Rev. General de Thermique, 1966, 5 (52), 349-359.

Godel explains the insights which led to the success of his Ignifluid stoker for the fluidized-bed combustion of low grade coal.

(9) Bishop, J. W., et al, Status of Direct Contact Heat Transfering Fluidized-Bed Boiler, American Society of Mechanical Engineers, Winter Meeting, December, 1968, Paper 68-WA/FU-4.

Bishop outlines the development of the Multi-cell Fluidized-Bed Boiler.

(10) Jonke, A. A., et al, Reduction of Atmospheric Pollution by the Application of Fluidized-Bed Combustion, Annual Report, July 1968-June 1969, ANL/ES-CEM-1001.

In Figure 21 a reduction in NO content, in flue gas, is shown when fine limestone is added to a fluidized bed combustor. NO is shown to decline as $SO_2$ declines. (See 11 below).

(11) Hammons, G. A. and Skopp, A., A Regenerative Limestone Process for Fluidized Bed Combustion and Desulfurization, Final Report, February 28, 1971 for Process Control Engineering Program, Air Pollution Control Office by Esso Research and Engineering.

In Figure 25 a reduction in NO content in flue gas, is shown when a bed of coarse limestone becomes inactive for sulfur removal. NO is shown to decline as $SO_2$ increases.

(12) Fine, D. H., et al, The Importance of Nitrogen in Coal as a Source of Nitrogen Oxide Emission from Furnaces, Fuels Research Laboratory, Massachusetts Institute of Technology. Delivered 64th Annual AIChE meeting. Undated (received December 1971).

Fine notes that the experimental finding that NO emissions from a fluidized-bed combustor are higher than anticipated is "due to the intimate contact between the particles and entraining air stream" (p. 19). He may be right.

(13) Hoy, H. R. and Roberts, A. G., Fluidized Combustion of Coal at High Pressure. Paper delivered at 64th Annual Meeting of the American Institute of Chemical Engineers, 28 November 1971.

Hoy shows in Figure 5 the following relationship between excess air and (overall combustion efficiency) 10% (96%), 20% (98%), 35% (99%), 33% (99.5%). These data are believed to be at 2-3 fps, 5-10 atmospheres, 1500°F, 2-3 foot static bed depth.

(14) Demming, L. F., Trip report contained in Progress Report No. 3, under Office of Coal Research Contract 14-01-0001-478, for May 1965.

Demming reports about his meeting at the Marchwood Laboratories of the British Central Electricity Generating Board. CEGB had found that operation of a fluidized bed combustor was good in a bed 3 ft deep with a superficial gas velocity of 0.5 fps. At 1 fps carryover becomes objectionable.

(15) The Federal Research and Development Plan for Air Pollution Control By Combustion-Process Modification, Final Report, January 11, 1971 for Air Pollution Control Office, Environmental Protection Agency by Battelle Memorial Institute, Columbus Laboratories.

(16) Rice, R. L., and Coates, N. H. Fluid Bed Combustion: Suitability of Coals and Bed Materials, Power Engineering, December 1971, 36-38.

Rice and Coates point out that in their apparatus lignite is the only coal which can be burned completely without recycle.

(17) U. S. Patent No. 3,171,369 (March 2, 1965).

Stephens and Goldberger show that high temperatures and high excess air are required for complete combustion of carbon in finely ground coal.

(18) U. S. Patents 2,818,049; 2,842,102; 2,976,853; 2,983,259; 2,997,031; 3,048,153 and 3,119,378.

First U. S. Patents for fluidized-bed boiler based on experimental work; concerned mainly with heterogenous combustion catalysts.

(19) Ehrlich, S., The Fluidized-Bed Boiler, paper presented at the 1968 Technical Meeting, The Eastern Section of the Combustion Institute, Amherst, Massachusetts, Cctober 22, 1968.

Describes problem areas in which basic research on fluidized bed combustion and heat transfer are required.

(20) Essenhigh, R. F., Professor - Department of Fuel Science, Pennsylvania State University. Various personal communications, February 1967 through November 1968.

Dr. Essenhigh describes areas in which useful and interesting basic research might be carried out in the field of fluidized bed combustion and heat transfer.

(21) American Society of Mechanical Engineers, Power Test Codes, Steam Generating Units PTC 4.1 - 1964 and Addenda.

This code presents the accepted accounting method for defining the efficiency of a boiler. PTC 4.1 - 1964, Section 9.4, recommends the use of 14,500 Btu/lb for the heating value of carbon in refuse. Note that this report has used 14,100 Btu/lb of "loss-on-ignition" which may or may not be all carbon.

(22) Essenhigh, R. F. and Fells, I., Discussions Faraday Society, Volume 31 (1961) page 208.

(23) Second International Conference on Fluidized Bed Combustion sponsored by Division of Process Control Engineering, National Air Pollution Control Administration at Hueston Woods Lodge, College Corner, Ohio on October 4-7, 1970.

Figure 15 of paper by Anastasia, et al, shows an optimum sulfur capture temperature of 1450°F to 1550°F (oxidizing conditions).

Figure 2 of paper by Moss shows optimum sulfur capture temperature of 875°C (oxidizing conditions).

(24) Goldberger, W. M., Collection of Fly Ash in a Self-Agglomerating Fluidized Bed Coal Burner, American Society of Mechanical Engineers, Winter Meeting, November, 1967, paper 67-WA/FU-3.

(25) Loshe, G. E. and Commander, R. E., Initial Operation of the Idaho Waste Calcining Facility with Radioactive Feed, Proceedings of the Symposium on the Solidification and Longterm Storage of Highly Radioactive Wastes, Richland, Washington, February 1966.

# APPENDIX A. FBC SPECIFICATIONS

1. Air Supply

Two centrifugal fans in series for 300 cfm at 30" w.g. connected to a smooth 4" diameter conduit 20' long. Air flow is controlled with a gate valve and monitored with a venturi, pitot pressure, static pressure, and temperature measurements.

2. Plenum

Mild steel, 1/4" thickness, 21" x 18" x 12" outside dimensions with 8" diameter air inlet.

3. Water Column

Mild steel, 1/4" thickness, 24" x 20" x 36" outside dimensions with 16" x 12" x 36" inside dimensions.

A. Wall on inlet air side contains:

- a) One nominal 3" diameter pipe for lightoff burner.
- b) One nominal 1" diameter instrument port.

B. Left wall (facing air inlet) contains:

- a) One nominal 2" diameter pipe with valve for removal of bed material.
- b) Eight nominal 1" diameter instrument ports at various levels.
- c) One nominal 1" diameter water outlet.
- d) One nominal 2" diameter pressure relief port.

C. Right wall (facing air inlet) contains:

- a) One rectangular 2" x 1" coal feed port.
- b) One nominal 3/4" diameter cooling water inlet.

D. Wall opposite the air inlet contains:

Three nominal 1-1/2" diameter ports.

E. Insulating liner consisting of 1/4" ASTM 446 stainless steel extending from the grate to connection with hood (37-3/4"). Internal dimensions are 9-3/8" x 13-1/4". Kaowool insulation is placed in the nominal 1" annular space.

4. Air Distribution Grid

The grid contains 130 stainless steel air distribution buttons spaced on 1-1/4" centers, each containing eight drilled ports, .089" diameter. The air is discharged downward at an angle of 15° to the horizontal. With the insulating liner installed, 42 of the air distribution buttons are not in service.

5. Water-cooled Hood

The hood is a truncated pyramid 24" x 20" at the bottom and 17" x 17" at the top, with a height of 24" and a flue opening 12" diameter. Material is #10 gauge mild steel. One 4" diameter observation port is provided with 1" diameter water ports and a 2" diameter pressure relief port.

6. Flue System

From the FBC-1 hood, the flue system is run in 12" diameter #10 gauge steel pipe to the induced draft fan. From the fan the pipe is continued at 6" diameter, again #10 gauge steel. All connecting sections are welded. The induced draft fan may be bypassed.

7. Dust Collector

The collector contains two 8" diameter centrifugal collector units with a dust hopper, rotary feeder, and a valve for fly ash removal.

# LOCATION OF THERMOCOUPLES FOR FBC

| NO. | LOCATION |
|---|---|
| 1 | Forced draft fan air |
| 2 | Plenum air |
| 3 | Bed 1-1/2" |
| 4 | Bed 3" |
| 5 | Bed 7" |
| 6 | Bed 11" |
| 7 | Bed 15" |
| 8 | Flue gas in hood |
| 9 | Air distribution grid |
| 10 | Dust collector outlet |
| 11 | Gas sample line inlet |
| 12 | Flue gas exit |
| 13 | Cooling water in |
| 14 | Cooling water out-hood |
| 15 | Cooling water out-water walls |
| 16 | Isokinetic probe (above lab) |
| 17 | Sample gas discharge |
| 18 | $NO_x$ sampling line |
| 19 | Flue gas exit (over lab) |

## APPENDIX B. FBM SPECIFICATIONS

1. Air Supply

   One centrifugal fan at 2500 cfm at 50" w.g. connected to 12-inch square duct which expands to full width of plenum at inlet. Air is controlled by means of a damper and monitored by an orifice.

2. Plenum

   Mild steel, ¼" thickness, 72" x 20¼" x 12" inside dimensions with a 6' x 1' air inlet.

3. Boiler Construction

   a. Single 20" steam drum
   b. Dual 6" lower headers
   c. 2½" risers on 4" centers for side walls
   d. 4" downcomers (external)
   e. 5'4" distance from grid to uninsulated bottom of steam drum
   f. Combustion space = 53 $ft^3$
   g. Projected heating surface = 80 $ft^2$
   h. Average direct contact surface = 30 $ft^2$
   i. Boiler capacity = 5000 lb/hr excluding convection heat transfer; 7000 lb/hr including convection heat transfer
   j. 8.75 $ft^2$ of bed area
   k. Heat release rate: 800,000 to 1,200,000 Btu/ft·hr
   l. Pressure rating: 300 psi design, 200 psi normal operating

## APPENDIX B. FBM SPECIFICATIONS (Continued)

4. Air Distribution Grid

   The grid contains 815 stainless steel air distribution buttons spaced on 1¼" centers, each containing eight drilled ports, .089" diameter. The air is discharged downward at an angle of 15° to the horizontal.

5. The flue system is fitted first with a water-cooled tube array for temperature quenching, and a two-pass, 104 tube (1" x 6'), 600° air preheater; this is followed by a second water-cooled gas cooler dust collector, which exits to a 16" duct. The system is drawn by a 4000 cfm, 5" w.g. static pressure, induced draft fan.

6. Dust Collector and Fly Ash Reinjection

   The dust collector contains twelve 10-inch diameter centrifugal collector units with a dust hopper, a 4" Allen-Sherman-Hoff rotary feeder for fly-ash reinjection and a valve for fly-ash removal.

7. Coal Input

   700 - 900 lbs per hour

8. Thermocouple locations are listed for both the FBM and CBC in Appendix C.

# APPENDIX C. CBC SPECIFICATIONS

1. Air Supply

Shares air supply with FBM from centrifugal fan rated at 2500 cfm, 50" w.g.

Air is supplied via a 4" nominal, Schedule 40 pipe, and is controlled by a gate valve.

2. Plenum

Mild steel, 1/4" thickness. Approximate dimensions are 10" wide, 10" deep, and 18" high. The plenum rests between the FBM's cross-headers.

3. Column

A. ASTM 446, stainless steel, 1/4" thickness, 10-5/8" x 15-5/8" inside cross section 56" high.

B. Gas outlet - 8" ID diameter starting at 48" level.

C. Front wall contains 18" x 24" access plate.

D. Left wall (facing toward common wall with FBM contains

E. Right wall (facing toward common wall with FBM contains

F. Back, or common wall, was varied. Open area, achieved by cutting holes in a steel baffle, varied from 90 square inches to 2 square inches.

4. Air Distribution Grid

Grid contains 96 ASTM 304 stainless steel air distribution buttons spaced on 1-1/4" centers, each containing eight drilled ports, 0.089" diameter. The air is discharged downward at an angle of 15° to the horizontal.

5. Flue System

From the CBC exhaust, the gas is carried in an 8" pipe to a 4" dust collector unit with a dust hopper, rotary feeder, and a valve for recirculating fly ash or discharging it to waste.

# LOCATION OF THERMOCOUPLES IN FBM/CBC SYSTEM

| NO. | LOCATION* |
|---|---|
| 1 | Inlet air |
| 2 | FBM 1-1/2" bed, 9-1/2" |
| 3 | FBM 1-1/2" bed, 28-1/2" |
| 4 | FBM 1-1/2" bed, 45-1/2" |
| 5 | FBM 1-1/2" bed, 63-1/2" |
| 6 | FBM 9-1/2" bed, 9-1/2" |
| 7 | FBM 9-1/2" bed, 28-1/2" |
| 8 | FBM 9-1/2" bed, 45-1/2" |
| 9 | FBM 9-1/2" bed, 63-1/2" |
| 10 | FBM 29-1/2" bed, 9-1/2" |
| 11 | FBM 29-1/2" bed, 28-1/2" |
| 12 | FBM 29-1/2" bed, 45-1/2" |
| 13 | FBM 29-1/2" bed, 63-1/2" |
| 14 | CBC 3" bed |
| 15 | CBC 9-1/2" bed |
| 16 | CBC 25-1/2" bed |
| 17 | CBC 42-1/2" bed |
| 18 | CBC stack after cooling coil |
| 19 | H.V.T. below steam drum |
| 20 | H.V.T. above steam drum |

| | |
|---|---|
| 21 | H.V.T. after convection bank |
| 22 | H.V.T. after air preheater |
| 23 | H.V.T. after economizer |
| 24 | Ash reinjection air |

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[*] Location in FBM described as height above grid, distance from front of unit.

## APPENDIX D. SELECTED CARBON BURNUP RESULTS

### TABLE D-1

### DATA FROM TEST C-319-1

Bed temperature, T = 1980°F
Air Rate, A = 700 lb/hr
Fuel Rate, F = 61 lb/hr
Carbon Content, c = .65 lb carbon/lb fuel
Bed Height, H = 22 inches

Calculate from data and Equations 7 and 8
Carbon Rate, C = F(c) = 61 (.65) = 39.7 lb/hr
Inert Rate, I = F-C = 21.3 lb/hr

Using the above data, Equation 6 and coefficients listed in Table 9, compute the combustion efficiency (CE), carbon dioxide ($CO_2$) and residual oxygen (RO):

$$\begin{aligned} \text{CE, \%} = & -13.78 + 0.05193\ (1980) + 0.0462\ (700) \\ & + 0.3831\ (22) - 0.8737\ (39.7) - 0.1905\ (21.3) \\ & + 0.00270\ (39.7)\ (21.3) = 93.3 \end{aligned}$$

$$\begin{aligned} \text{CO}_2\text{, \%} = & -2.498 + 0.007547\ (1980) - 0.01190\ (700) \\ & + 0.1382\ (22) + 0.1589\ (39.7) + 0.01259\ (21.3) \\ & -0.0002862\ (39.7)\ (21.3) = 13.5 \end{aligned}$$

$$\begin{aligned} \text{RO, \%} = & \ 22.91 - 0.007353\ (1980) + 0.013\ (700) \\ & -0.1390\ (22) - 0.1769\ (39.7) - 0.01756\ (21.3) \\ & + 0.0003587\ (39.7)\ (21.3) = 7.3 \end{aligned}$$

## TABLE D-2

## CBC DATA FOR FBM TEST B-14

Bed Temperature = 2000°F
Air Rate = 1350 lb/hr
Fuel Rate = 139 lb/hr
Carbon Content = .48 lb carbon/lb fuel
Bed Height (static) = 14 in.

For the CBC with a plan area of 1.1 $ft^2$ the normalized data (normalized to unit plan area) are:

Air Rate, A' = 1227 lb/hr $ft^2$
Carbon Rate, C' = 60.6 lb/hr $ft^2$
Inert Rate, I' = 65.7 lb/hr $ft^2$

# APPENDIX E

## SUMMARY OF FBC DATA FOR CARBON-BURNUP CELL TESTS

## TABLE E-1. SUMMARY OF FBC DATA FOR CARBON-BURNUP CELL TESTS

| Test Number | C-302-1* | C-302-2* | C-302-3* | C-303-1* |
|---|---|---|---|---|
| **Reactor Data** | | | | |
| 1. Air rate, lb/hr | 760 | 760 | 820 | 330 |
| 2. Bed temperature, °F | 1800 | 2140 | 1780 | 1750 |
| 3. Starter bed | | | | |
| a. Static depth, in. | 18 | 18 | 18 | 18 |
| b. Sieve size, USS | -16+22 | | | |
| 4. Fly ash input, lb/hr (dry) | 73 | 105 | 72 | 41 |
| a. Water content, % (a.r.) | 12.6 | 10.0 | 4.3 | 26.9 |
| b. Carbon content, % dry | 54.9 | 50.5 | 60.5 | 50.9 |
| 5. Coal input, lb/hr | 0 | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- | -- |
| 6. Total carbon input, lb/hr | 40.4 | 53.5 | 43.7 | 20.8 |
| 7. Fly ash output, lb/hr | 37 | 62 | 35 | 28 |
| a. Carbon content, % | 11.2 | 15.5 | 17.2 | 28.3 |
| 8. Carbon burned, lb/hr | 36.4 | 44.0 | 37.8 | 13.4 |
| 9. Combustion efficiency, % | 90.1 | 82.2 | 86.5 | 64.4 |
| 10. Superficial velocity, fps | 14.6 | 17.0 | 15.6 | 5.7 |
| 11. Cooling probe position | not used | | | |
| **Flue Gas Data** | | | | |
| 12. $CO_2$, % | 11.5 | 14.2 | 10.8 | 9.8 |
| 13. $O_2$, % | 9.1 | 6.4 | 9.7 | 10.8 |
| 14. CO, % | 0 | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 350 | 800 | 550 | 320 |
| 16. NO, ppm | 560 | 670 | 570 | 470 |
| 17. Hydrocarbons, ppm | Negligible | | | |
| 18. Particulate emission | No data | | | |
| a. To stack, lb/hr | | | | |
| b. Carbon content, % | | | | |
| **Heat Balance Data** | | | | |
| 19. Losses, kBtu/hr | | | | |
| a. Probe | 0 | 0 | 0 | 0 |
| b. Wall | 40 | 44 | 27 | 36 |
| c. Hood | 36 | 60 | 45 | 34 |
| d. Gas | 390 | 478 | 401 | 182 |
| e. Dust | 14 | 28 | 13 | 10 |
| f. Total | 480 | 610 | 486 | 262 |
| 20. Input, kBtu/hr From 8. | 513 | 620 | 533 | 189 |

*Test results are suspect. Presented for record purposes only.

TABLE E-1. (Continued)

| Test Number | C-303-2* | C-303-3* | C-303-4* | C-303-5* |
|---|---|---|---|---|
| **Reactor Data** | | | | |
| 1. Air rate, lb/hr | 330 | 500 | 500 | 500 |
| 2. Bed temperature, °F | 1980 | 1750 | 1860 | 1970 |
| 3. Starter bed | | | | |
| a. Static depth, in. | 18 | 18 | 18 | 18 |
| b. Sieve size, USS | ..............-16+22........... | | | |
| 4. Fly ash input, lb/hr (dry) | 46 | 68 | 67 | 71 |
| a. Water content, % (a.r.) | 24.9 | 26.4 | 25.2 | 26.0 |
| b. Carbon content, % dry | 52.5 | 57.0 | 55.5 | 54.5 |
| 5. Coal input, lb/hr | 0 | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- | -- |
| 6. Total carbon input, lb/hr | 24.1 | 38.7 | 37.0 | 30.7 |
| 7. Fly ash output, lb/hr | 30 | 43 | 39 | 41 |
| a. Carbon content, % | 27.5 | 32.0 | 24.4 | 20.9 |
| 8. Carbon burned, lb/hr | 15.8 | 25.0 | 27.3 | 30.1 |
| 9. Combustion efficiency, % | 65.6 | 64.6 | 73.8 | 77.8 |
| 10. Superficial velocity, fps | 6.9 | 9.4 | 9.9 | 10.5 |
| 11. Cooling probe position | .............Not used............ | | | |
| **Flue Gas Data** | | | | |
| 12. $CO_2$, % | 11.5 | 12.0 | 13.2 | 14.5 |
| 13. $O_2$, % | 9.0 | 8.5 | 7.4 | 6.2 |
| 14. CO, % | 0 | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 600 | 320 | 500 | 700 |
| 16. NO, ppm | 480 | 410 | 570 | 600 |
| 17. Hydrocarbons, ppm | ...........Negligible............ | | | |
| 18. Particulate emission | ............No data.............. | | | |
| a. To stack, lb/hr | | | | |
| b. Carbon content, % | | | | |
| **Heat Balance Data** | | | | |
| 19. Losses, kBtu/hr | | | | |
| a. Probe | 0 | 0 | 0 | 0 |
| b. Wall | 27 | 36 | 43 | 50 |
| c. Hood | 36 | 36 | 41 | 17 |
| d. Gas | 193 | 281 | 299 | 322 |
| e. Dust | 13 | 16 | 15 | 17 |
| f. Total | 269 | 369 | 398 | 406 |
| 20. Input, kBtu/hr | | | | |
| From 8. | 223 | 353 | 385 | 424 |

*Tests results are suspect. Presented for record purposes only.

TABLE E-1. (Continued)

| Test Number | C-306-1 | C-306-2 | C-306-3 | C-306-4 |
|---|---|---|---|---|
| **Reactor Data** | | | | |
| 1. Air rate, lb/hr | 600 | 860 | 775 | 600 |
| 2. Bed temperature, °F | 2100 | 1750 | 2120 | 1800 |
| 3. Starter bed | | | | |
| a. Static depth, in. | 18 | 18 | 18 | 18 |
| b. Sieve size, USS | -8+16 | | | |
| 4. Fly ash input, lb/hr (dry) | 106 | 134 | 130 | 105 |
| a. Water content, % (a.r.) | 0.4 | 0.6 | 0.5 | 0.5 |
| b. Carbon content, % dry | 39.2 | 41.7 | 40.3 | 40.9 |
| 5. Coal input, lb/hr | 0 | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- | -- |
| 6. Total carbon input, lb/hr | 41.5 | 56.0 | 52.5 | 43.0 |
| 7. Fly ash output, lb/hr | 70 | 95 | 82 | 74 |
| a. Carbon content, % | 8.5 | 17.5 | 7.5 | 16.0 |
| 8. Carbon burned, lb/hr | 35.5 | 39.2 | 46.5 | 31.2 |
| 9. Combustion efficiency, % | 85.5 | 70.0 | 88.7 | 72.7 |
| 10. Superficial velocity, fps | 11.8 | 14.5 | 15.2 | 10.5 |
| 11. Cooling probe position | Not used | | | |
| **Flue Gas Data** | | | | |
| 12. $CO_2$, % | 14.3 | 11.0 | 14.6 | 12.4 |
| 13. $O_2$, % | 6.2 | 9.3 | 5.8 | 8.2 |
| 14. CO, % | 0 | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 800 | 750 | 800 | 750 |
| 16. NO, ppm | 720 | 560 | 770 | 520 |
| 17. Hydrocarbons, ppm | Negligible | | | |
| 18. Particulate emission | | | | |
| a. To stack, lb/hr | 1.27 | 4.15 | 1.71 | 2.97 |
| b. Carbon content, % | 9.0 | 10.5 | 9.3 | 5.3 |
| **Heat Balance Data** | | | | |
| 19. Losses, kBtu/hr | | | | |
| a. Probe | 0 | 0 | 0 | 0 |
| b. Wall | 38 | 32 | 33 | 36 |
| c. Hood | 72 | 68 | 79 | 75 |
| d. Gas | 354 | 406 | 461 | 294 |
| e. Dust | 31 | 35 | 37 | 28 |
| f. Total | 495 | 541 | 610 | 433 |
| 20. Input, kBtu/hr From 8. | 501 | 553 | 656 | 440 |

TABLE E-1. (Continued)

| Test Number | C-307-1 | C-307-2 | C-307-3 | C-307-4 |
|---|---|---|---|---|
| **Reactor Data** | | | | |
| 1. Air rate, lb/hr | 600 | 330 | 600 | 330 |
| 2. Bed temperature, °F | 1800 | 2100 | 2140 | 1750 |
| 3. Starter bed | | | | |
| a. Static depth, in. | 17 | 18 | 18 | 18 |
| b. Sieve size, USS | -16+22 | | | |
| 4. Fly ash input, lb/hr (dry) | 106 | 65 | 110 | 73 |
| a. Water content, % (a.r.) | 0.2 | 0.4 | 0.1 | 0.2 |
| b. Carbon content, % dry | 39.2 | 39.5 | 40.0 | 40.3 |
| 5. Coal input, lb/hr | 0 | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- | -- |
| 6. Total carbon input, lb/hr | 40.6 | 25.7 | 44.0 | 29.4 |
| 7. Fly ash output, lb/hr | 81 | 42 | 70 | 54 |
| a. Carbon content, % | 11.9 | 6.0 | 6.2 | 19.6 |
| 8. Carbon burned, lb/hr | 30.9 | 23.2 | 39.7 | 18.7 |
| 9. Combustion efficiency, % | 76.0 | 90.0 | 91.0 | 63.0 |
| 10. Superficial velocity, fps | 9.8 | 6.6 | 11.9 | 5.7 |
| 11. Cooling probe position | Not used | | | |
| **Flue Gas Data** | | | | |
| 12. $CO_2$, % | 13.2 | 16.8 | 16.0 | 13.4 |
| 13. $O_2$, % | 7.5 | 3.5 | 4.8 | 6.9 |
| 14. CO, % | 0 | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 500 | 600 | 650 | 700 |
| 16. NO, ppm | 410 | 500 | 550 | 370 |
| 17. Hydrocarbons, ppm | 0 | 0 | 0 | 0 |
| 18. Particulate emission | | | | |
| a. To stack, lb/hr | 3.0 | 0.4 | 1.5 | 4.1 |
| b. Carbon content, % | 8.3 | -- | 4.7 | 9.4 |
| **Heat Balance Data** | | | | |
| 19. Losses, kBtu/hr | | | | |
| a. Probe | 0 | 0 | 0 | 0 |
| b. Wall | 50 | 48 | 47 | 32 |
| c. Hood | 53 | 55 | 68 | 58 |
| d. Gas | 293 | 197 | 361 | 157 |
| e. Dust | 31 | 19 | 32 | 20 |
| f. Total | 427 | 319 | 508 | 267 |
| 20. Input, kBtu/hr From 8. | 436 | 327 | 560 | 264 |

TABLE E-1. (Continued)

| Test Number | C-308-1 | C-308-2 | C-308-3 |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 600 | 600 | 600 |
| 2. Bed temperature, °F | 1900[1] | 1900[2] | 1900[3] |
| 3. Starter bed | | | |
| a. Static depth, in. | 18 | 18 | 18 |
| b. Sieve size, USS | ............-16+22 .... | | |
| 4. Fly ash input, lb/hr (dry) | 87 | 100 | 102 |
| a. Water content, % (a.r.) | 0 | 0 | 0 |
| b. Carbon content, % dry | 43.5 | 42.4 | 42.2 |
| 5. Coal input, lb/hr | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- |
| 6. Total carbon input, lb/hr | 37.8 | 42.4 | 43.0 |
| 7. Fly ash output, lb/hr | 56 | 66 | 68 |
| a. Carbon content, % | 12.0 | 12.3 | 12.4 |
| 8. Carbon burned, lb/hr | 31.0 | 34.3 | 34.6 |
| 9. Combustion efficiency, % | 81.5 | 81.3 | 80.6 |
| 10. Superficial velocity, fps | 10.2 | 10.3 | 10.2 |
| 11. Cooling probe position | ........Not used........ | | |
| **Flue Gas Data** | | | |
| 12. $CO_2$, % | 13.4 | 14.5 | 15.0 |
| 13. $O_2$, % | 7.1 | 6.2 | 5.5 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 600 | 750 | 800 |
| 16. NO, ppm | 520 | 500 | 520 |
| 17. Hydrocarbons, ppm | 30 | 30 | 30 |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | 2.7 | 2.5 | 3.1 |
| b. Carbon content, % | 8.5 | 8.6 | 9.4 |
| **Heat Balance Data** | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 0 | 0 | 0 |
| b. Wall | 55 | 53 | 53 |
| c. Hood | 69 | 90 | 92 |
| d. Gas | 310 | 311 | 311 |
| e. Dust | 23 | 26 | 27 |
| f. Total | 457 | 480 | 483 |
| 20. Input, kBtu/hr From 8. | 437 | 484 | 488 |

Notes: [1] Temperature stabilized with slow rise to 1900°F

[2] After temperature held at 2100°F for 15 min.

[3] After temperature held at 1500°F for 15 min.

TABLE E-1. (Continued)

| Test Number | C-310-1 | C-310-2 |
|---|---|---|
| Reactor Data | | |
| 1. Air rate, lb/hr | 700 | 700 |
| 2. Bed temperature, °F | 2040 | 2000 |
| 3. Starter bed | | |
| a. Static depth, in. | 18 | 18 |
| b. Sieve size, USS | ...-16+22.... | |
| 4. Fly ash input, lb/hr (dry) | 105 | 125 |
| a. Water content, % (a.r.) | 0.5 | 0.5 |
| b. Carbon content, % dry | 44.0 | 44.0 |
| 5. Coal input, lb/hr | 0 | 0 |
| a. Carbon content, % | -- | -- |
| 6. Total carbon input, lb/hr | 46.2 | 55.0 |
| 7. Fly ash output, lb/hr | 65 | 83 |
| a. Carbon content, % | 11.0 | 16.0 |
| 8. Carbon burned, lb/hr | 39.2 | 41.7 |
| 9. Combustion efficiency, % | 84.6 | 75.8 |
| 10. Superficial velocity, fps | 12.9 | 13.2 |
| 11. Cooling probe position | OUT | IN |
| Flue Gas Data | | |
| 12. $CO_2$, % | 13.2 | 13.8 |
| 13. $O_2$, % | 7.1 | 6.4 |
| 14. CO, % | 0 | 0 |
| 15. $SO_2$, ppm | 700 | 720 |
| 16. NO, ppm | 460 | 420 |
| 17. Hydrocarbons, ppm | ...Negligible... | |
| 18. Particulate emission | | |
| a. To stack, lb/hr | 0.88 | 1.10 |
| b. Carbon content, % | 13.7 | 15.2 |
| Heat Balance Data | | |
| 19. Losses, kBtu/hr | | |
| a. Probe | 6 | 67 |
| b. Wall | 55 | 57 |
| c. Hood | 30 | 32 |
| d. Gas | 397 | 390 |
| e. Dust | 28 | 35 |
| f. Total | 516 | 581 |
| 20. Input, kBtu/hr From 8. | 553 | 588 |

TABLE E-1. (Continued)

| Test Number | C-311-1 | C-311-2 | C-311-3 |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 700 | 700 | 600 |
| 2. Bed temperature, °F | 2000 | 1930 | 2000 |
| 3. Starter bed | | | |
| a. Static depth, in. | 18 | 18 | 18 |
| b. Sieve size, USS | ........-16+22......... | | |
| 4. Fly ash input, lb/hr (dry) | 158 | 190 | 250 |
| a. Water content, % (a.r.) | 1.2 | 2.0 | 1.5 |
| b. Carbon content, % dry | 33.0 | 33.0 | 31.3 |
| 5. Coal input, lb/hr | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- |
| 6. Total carbon input, lb/hr | 52.0 | 63.0 | 79.5 |
| 7. Fly ash output, lb/hr | 118 | 148 | 205 |
| a. Carbon content, % | 10.2 | 14.2 | 16.8 |
| 8. Carbon burned, lb/hr | 40.0 | 42.0 | 45.0 |
| 9. Combustion efficiency, % | 77.0 | 66.7 | 58.0 |
| 10. Superficial velocity, fps | 11.3 | 11.1 | 11.1 |
| 11. Cooling probe position | OUT | IN | IN |
| **Flue Gas Data** | | | |
| 12. $CO_2$, % | 16.0 | 16.6 | 18.1 |
| 13. $O_2$, % | 4.5 | 3.6 | 2.1 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 700 | 800 | 850 |
| 16. NO, ppm | 570 | 470 | 470 |
| 17. Hydrocarbons, ppm | 0 | 50 | 0 |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | 1.32 | 2.78 | 7.5 |
| b. Carbon content, % | 9.1 | 14.0 | 15.5 |
| **Heat Balance Data** | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 14 | 68 | 76 |
| b. Wall | 55 | 68 | 68 |
| c. Hood | 32 | 32 | 33 |
| d. Gas | 393 | 379 | 346 |
| e. Dust | 50 | 60 | 87 |
| f. Total | 544 | 607 | 610 |
| 20. Input, kBtu/hr From 8. | 564 | 592 | 635 |

TABLE E-1. (Continued)

| Test Number | C-311-4 | C-311-5 | C-311-6 |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 600 | 800 | 800 |
| 2. Bed temperature, °F | 1980 | 1950 | 1800* |
| 3. Starter bed | | | |
| a. Static depth, in. | 18 | 18 | 18 |
| b. Sieve size, USS | -16+22 | | |
| 4. Fly ash input, lb/hr (dry) | 158 | 232 | 302 |
| a. Water content, % (a.r.) | 1.5 | 1.2 | 1.8 |
| b. Carbon content, % dry | 32.0 | 32.0 | 32.0 |
| 5. Coal input, lb/hr | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- |
| 6. Total carbon input, lb/hr | 50.5 | 74.5 | 96.0 |
| 7. Fly ash output, lb/hr | 120 | 186 | 250 |
| a. Carbon content, % | 10.5 | 13.9 | 16.9 |
| 8. Carbon burned, lb/hr | 38.0 | 48.2 | 52.0 |
| 9. Combustion efficiency, % | 75.8 | 64.5 | 56.2 |
| 10. Superficial velocity, fps | 10.5 | 13.7 | 13.2 |
| 11. Cooling probe position | OUT | OUT | IN |
| **Flue Gas Data** | | | |
| 12. $CO_2$, % | 15.9 | 15.8 | 16.4 |
| 13. $O_2$, % | 4.5 | 4.7 | 3.9 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 820 | Inst. Failure | |
| 16. NO, ppm | 500 | 510 | 350 |
| 17. Hydrocarbons, ppm | Negl. | Inst. Failure | |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | 1.77 | 5.3 | 7.1 |
| b. Carbon content, % | 12.3 | 17.4 | 15.6 |
| **Heat Balance Data** | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 16 | 17 | 67 |
| b. Wall | 68 | 68 | 65 |
| c. Hood | 45 | 65 | 68 |
| d. Gas | 334 | 438 | 406 |
| e. Dust | 50 | 77 | 95 |
| f. Total | 513 | 665 | 701 |
| 20. Input, kBtu/hr From 8. | 536 | 680 | 733 |

---

*Declining temperature

TABLE E-1. (Continued)

| Test Number | C-312-1 | C-312-2 | C-312-3 |
|---|---|---|---|
| Reactor Data | | | |
| 1. Air rate, lb/hr | 700 | 700 | 700 |
| 2. Bed temperature, °F | 2000 | 1990 | 1980 |
| 3. Starter bed | | | |
| a. Static depth, in. | 18 | 14 | 10 |
| b. Sieve size, USS | ........-16+22........ | | |
| 4. Fly ash input, lb/hr (dry) | 140 | 138 | 156 |
| a. Water content, % (a.r.) | 0.2 | 0.3 | 0.2 |
| b. Carbon content, % dry | 33.0 | 33.0 | 32.8 |
| 5. Coal input, lb/hr | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- |
| 6. Total carbon input, lb/hr | 46.5 | 45.7 | 51.3 |
| 7. Fly ash output, lb/hr | 97 | 96 | 117 |
| a. Carbon content, % | 7.5 | 7.7 | 10.8 |
| 8. Carbon burned, lb/hr | 39.2 | 38.3 | 37.4 |
| 9. Combustion efficiency, % | 84.2 | 83.8 | 76.0 |
| 10. Superficial velocity, fps | 13.6 | 13.6 | 13.5 |
| 11. Cooling probe position | ........ Not used ...... | | |
| Flue Gas Data | | | |
| 12. $CO_2$, % | 13.8 | 13.5 | 13.2 |
| 13. $O_2$, % | 7.0 | 7.2 | 7.5 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 900 | 900 | 900 |
| 16. NO, ppm | 500 | 480 | 480 |
| 17. Hydrocarbons, ppm | ...... Negligible ...... | | |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | 2.2 | 2.3 | 1.5 |
| b. Carbon content, % | -- | -- | -- |
| Heat Balance Data | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 0 | 0 | 0 |
| b. Wall | 67 | 58 | 47 |
| c. Hood | 27 | 28 | 29 |
| d. Gas | 389 | 385 | 383 |
| e. Dust | 41 | 40 | 49 |
| f. Total | 524 | 511 | 508 |
| 20. Input, kBtu/hr From 8. | 553 | 540 | 527 |

TABLE E-1. (Continued)

| Test Number | C-313-1 | C-313-2 | C-313-3 |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 600 | 700 | 800 |
| 2. Bed temperature, °F | 1980 | 1980 | 1980 |
| 3. Starter bed | | | |
| a. Static depth, in. | 18 | 18 | 18 |
| b. Sieve size, USS | ........-16+22........ | | |
| 4. Fly ash input, lb/hr (dry) | 63 | 74 | 84 |
| a. Water content, % (a.r.) | 0 | 0 | 0 |
| b. Carbon content, % dry | 58.0 | 58.0 | 57.5 |
| 5. Coal input, lb/hr | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- |
| 6. Total carbon input, lb/hr | 36.6 | 43.2 | 48.2 |
| 7. Fly ash output, lb/hr | 29.6 | 36.5 | 45.2 |
| a. Carbon content, % | 11.5 | 16.0 | 19.0 |
| 8. Carbon burned, lb/hr | 33.4 | 37.5 | 41.5 |
| 9. Combustion efficiency, % | 91.0 | 87.2 | 86.0 |
| 10. Superficial velocity, fps | 11.5 | 13.5 | 15.2 |
| 11. Cooling probe position | .......Not used......... | | |
| **Flue Gas Data** | | | |
| 12. $CO_2$, % | 13.3 | 12.9 | 12.8 |
| 13. $O_2$, % | 7.5 | 8.0 | 8.1 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 1160 | 1160 | 1160 |
| 16. NO, ppm | 630 | 610 | 600 |
| 17. Hydrocarbons, ppm | .......Negligible....... | | |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | 1.4 | 1.0 | 1.7 |
| b. Carbon content, % | 5.7 | -- | 6.2 |
| **Heat Balance Data** | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 0 | 0 | 0 |
| b. Wall | 75 | 84 | 89 |
| c. Hood | 25 | 33 | 33 |
| d. Gas | 328 | 382 | 436 |
| e. Dust | 12 | 15 | 19 |
| f. Total | 440 | 514 | 577 |
| 20. Input, kBtu/hr From 8. | 471 | 529 | 585 |

TABLE E-1. (Continued)

| Test Number | C-314-1 | C-314-2 | C-314-3 | C-314-4* |
|---|---|---|---|---|
| **Reactor Data** | | | | |
| 1. Air rate, lb/hr | 600 | 700 | 800 | 700 |
| 2. Bed temperature, °F | 1980 | 1960 | 1980 | 1980 |
| 3. Starter bed | | | | |
| a. Static depth, in. | 10 | 10 | 10 | 10 |
| b. Sieve size, USS | -16+22 | | | |
| 4. Fly ash input, lb/hr (dry) | 70 | 82 | 102 | 53 |
| a. Water content, % (a.r.) | 0 | 0 | 0 | 0 |
| b. Carbon content, % dry | 56.0 | 56.0 | 56.0 | 56.0 |
| 5. Coal input, lb/hr | 0 | 0 | 0 | 15 |
| a. Carbon content, % | -- | -- | -- | 64.0 |
| 6. Total carbon input, lb/hr | 39.2 | 45.8 | 57.0 | 39.6 |
| 7. Fly ash output, lb/hr | 36.6 | 44.5 | 58.3 | 29.6 |
| a. Carbon content, % | 16.5 | 18.6 | 25.5 | 14.6 |
| 8. Carbon burned, lb/hr | 33.2 | 37.5 | 43.3 | 35.4 |
| 9. Combustion efficiency, % | 84.8 | 82.0 | 76.0 | 89.5 |
| 10. Superficial velocity, fps | 11.5 | 13.3 | 15.3 | 13.4 |
| 11. Cooling probe position | Not used | | | |
| **Flue Gas Data** | | | | |
| 12. $CO_2$, % | 13.4 | 13.0 | 13.0 | 12.2 |
| 13. $O_2$, % | 7.4 | 7.6 | 7.8 | 8.5 |
| 14. CO, % | 0 | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 1050 | 1050 | 1050 | 1500 |
| 16. NO, ppm | 560 | 560 | 560 | 530 |
| 17. Hydrocarbons, ppm | Negligible | | | |
| 18. Particulate emission | | | | |
| a. To stack, lb/hr | 0.9 | 0.5 | 0.6 | 0.5 |
| b. Carbon content, % | | 18.6 | | |
| **Heat Balance Data** | | | | |
| 19. Losses, kBtu/hr | | | | |
| a. Probe | 0 | 0 | 0 | 0 |
| b. Wall | 84 | 85 | 88 | 78 |
| c. Hood | 27 | 29 | 31 | 27 |
| d. Gas | 328 | 378 | 436 | 394 |
| e. Dust | 15 | 18 | 24 | 12 |
| f. Total | 454 | 510 | 579 | 511 |
| 20. Input, kBtu/hr From 8. | 468 | 529 | 611 | 499 |

*Heat balance data adjusted for hydrogen content of coal.

TABLE E-1. (Continued)

| Test Number | C-315-1 | C-315-2 | C-315-3 |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 600 | 700 | 800 |
| 2. Bed temperature, °F | 2010 | 2000 | 1980 |
| 3. Starter bed | | | |
| a. Static depth, in. | 18 | 18 | 18 |
| b. Sieve size, USS | ........-16+22........ | | |
| 4. Fly ash input, lb/hr (dry) | 81 | 86 | 92 |
| a. Water content, % (a.r.) | 0 | 0 | 0 |
| b. Carbon content, % dry | 57.5 | 57.0 | 56.0 |
| 5. Coal input, lb/hr | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- |
| 6. Total carbon input, lb/hr | 46.3 | 48.8 | 51.8 |
| 7. Fly ash output, lb/hr | 42.4 | 44.8 | 47.8 |
| a. Carbon content, % | 18.2 | 17.0 | 16.0 |
| 8. Carbon burned, lb/hr | 38.6 | 41.2 | 44.2 |
| 9. Combustion efficiency, % | 83.3 | 84.5 | 85.2 |
| 10. Superficial velocity, fps | 11.7 | 13.5 | 15.3 |
| 11. Cooling probe position | IN | IN | IN |
| **Flue Gas Data** | | | |
| 12. $CO_2$, % | 15.4 | 15.0 | 14.6 |
| 13. $O_2$, % | 5.5 | 5.8 | 6.0 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 1250 | 1250 | 1250 |
| 16. NO, ppm | 560 | 560 | 540 |
| 17. Hydrocarbons, ppm | .......Negligible....... | | |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | 0.7 | 0.7 | 0.8 |
| b. Carbon content, % | 5.3 | 7.4 | 8.2 |
| **Heat Balance Data** | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 31 | 31 | 31 |
| b. Wall | 100 | 113 | 115 |
| c. Hood | 22 | 24 | 26 |
| d. Gas | 337 | 389 | 437 |
| e. Dust | 18 | 19 | 20 |
| f. Total | 508 | 576 | 629 |
| 20. Input, kBtu/hr | | | |
| From 8. | 544 | 581 | 623 |

TABLE E-1. (Continued)

| Test Number | C-316-1 | C-316-2 | C-316-3 |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 600 | 700 | 800 |
| 2. Bed temperature, °F | 2000 | 1990 | 1980 |
| 3. Starter bed | | | |
| a. Static depth, in. | 10 | 10 | 10 |
| b. Sieve size, USS | ........-16+22........ | | |
| 4. Fly ash input, lb/hr (dry) | 86 | 94 | 109 |
| a. Water content, % (a.r.) | 0 | 0 | 0 |
| b. Carbon content, % dry | 58.0 | 58.0 | 58.0 |
| 5. Coal input, lb/hr | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- |
| 6. Total carbon input, lb/hr | 49.6 | 54.4 | 63.5 |
| 7. Fly ash output, lb/hr | 49.1 | 52.4 | 63.3 |
| a. Carbon content, % | 26.3 | 24.4 | 28.1 |
| 8. Carbon burned, lb/hr | 36.9 | 41.6 | 45.7 |
| 9. Combustion efficiency, % | 74.5 | 76.5 | 72.0 |
| 10. Superficial velocity, fps | 11.6 | 13.5 | 15.3 |
| 11. Cooling probe position | IN | IN | IN |
| **Flue Gas Data** | | | |
| 12. $CO_2$, % | 14.8 | 14.4 | 13.6 |
| 13. $O_2$, % | 5.8 | 6.0 | 6.4 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 1250 | 1300 | 1250 |
| 16. NO, ppm | 520 | 510 | 520 |
| 17. Hydrocarbons, ppm | .......Negligible....... | | |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | 0.9 | 0.8 | 0.9 |
| b. Carbon content, % | 4.0 | 5.4 | 6.2 |
| **Heat Balance Data** | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 31 | 31 | 31 |
| b. Wall | 92 | 99 | 100 |
| c. Hood | 26 | 28 | 29 |
| d. Gas | 334 | 386 | 438 |
| e. Dust | 21 | 22 | 26 |
| f. Total | 504 | 576 | 624 |
| 20. Input, kBtu/hr<br>From 8. | 520 | 587 | 644 |

TABLE E-1. (Continued)

| Test Number | C-317-1 | C-317-2 | C-318-1 | C-318-2 |
|---|---|---|---|---|
| **Reactor Data** | | | | |
| 1. Air rate, lb/hr | 700 | 700 | 700 | 700 |
| 2. Bed temperature, °F | 1980 | 1980 | 1900 | 1950 |
| 3. Starter bed | | | | |
| a. Static depth, in. | 22 | 18 | 22 | 18 |
| b. Sieve size, USS | ....-8+16.... | | ....-16+22... | |
| 4. Fly ash input, lb/hr (dry) | 95 | 94 | 104 | 97 |
| a. Water content, % (a.r.) | 0 | 0 | 0 | 0 |
| b. Carbon content, % dry | 54.4 | 54.4 | 58.6 | 58.2 |
| 5. Coal input, lb/hr | 0 | 0 | 0 | 0 |
| a. Carbon content, % | -- | -- | -- | -- |
| 6. Total carbon input, lb/hr | 51.7 | 51.4 | 61.0 | 56.0 |
| 7. Fly ash output, lb/hr | 51.0 | 51.3 | 60.8 | 52.0 |
| a. Carbon content, % | 18.0 | 18.8 | 28.4 | 21.8 |
| 8. Carbon burned, lb/hr | 42.5 | 41.7 | 44.0 | 44.6 |
| 9. Combustion efficiency, % | 82.2 | 81.5 | 72.5 | 79.8 |
| 10. Superficial velocity, fps | 13.5 | 13.5 | 13.0 | 13.3 |
| 11. Cooling probe position | ...........Not used............. | | | |
| **Flue Gas Data** | | | | |
| 12. $CO_2$, % | 15.5 | 15.3 | 15.8 | 16.0 |
| 13. $O_2$, % | 4.8 | 5.3 | 5.0 | 4.5 |
| 14. CO, % | 0 | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 1200 | 1275 | 1050 | 1050 |
| 16 NO, ppm | 570 | 570 | 550 | 550 |
| 17. Hydrocarbons, ppm | 15 | 15 | 20 | 20 |
| 18. Particulate emission | | | | |
| a. To stack, lb/hr | 0.68 | 0.60 | 0.72 | 0.75 |
| b. Carbon content, % | 7.4 | 6.1 | 9.7 | 5.7 |
| **Heat Balance Data** | | | | |
| 19. Losses, kBtu/hr | | | | |
| a. Probe | 0 | 0 | 0 | 0 |
| b. Wall | 129 | 123 | 172 | 150 |
| c. Hood | 19 | 20 | 22 | 27 |
| d. Gas | 385 | 384 | 365 | 380 |
| e. Dust | 21 | 21 | 24 | 21 |
| f. Total | 554 | 548 | 583 | 578 |
| 20. Input, kBtu/hr From 8. | 599 | 588 | 620 | 629 |

TABLE E-1. (Continued)

| Test Number | C-319-1 | C-319-2 |
|---|---|---|
| **Reactor Data** | | |
| 1. Air rate, lb/hr | 700 | 700 |
| 2. Bed temperature, °F | 1980 | 2000 |
| 3. Starter bed | | |
| a. Static depth, in. | 22 | 18 |
| b. Sieve size, USS | ....-8+16... | |
| 4. Fly ash input, lb/hr (dry) | 61 | 59 |
| a. Water content, % (a.r.) | 0 | 0 |
| b. Carbon content, % dry | 65.0 | 65.0 |
| 5. Coal input, lb/hr | 0 | 0 |
| a. Carbon content, % | -- | -- |
| 6. Total carbon input, lb/hr | 39.8 | 38.4 |
| 7. Fly ash output, lb/hr | 26.2 | 24.3 |
| a. Carbon content, % | 17.0 | 15.0 |
| 8. Carbon burned, lb/hr | 35.3 | 34.8 |
| 9. Combustion efficiency, % | 89.0 | 90.6 |
| 10. Superficial velocity, fps | 13.5 | 13.5 |
| 11. Cooling probe position | ...Not used.... | |
| **Flue Gas Data** | | |
| 12. $CO_2$, % | 11.8 | 11.8 |
| 13. $O_2$, % | 8.2 | 8.7 |
| 14. CO, % | 0 | 0 |
| 15. $SO_2$, ppm | 1150 | 1100 |
| 16. NO, ppm | 520 | 575 |
| 17. Hydrocarbons, ppm | 12 | 12 |
| 18. Particulate emission | | |
| a. To stack, lb/hr | 0.60 | 0.61 |
| b. Carbon content, % | 6.5 | 5.5 |
| **Heat Balance Data** | | |
| 19. Losses, kBtu/hr | | |
| a. Probe | 0 | 0 |
| b. Wall | 47 | 50 |
| c. Hood | 13 | 15 |
| d. Gas | 381 | 386 |
| e. Dust | 11 | 10 |
| f. Total | 452 | 461 |
| 20. Input, kBtu/hr<br>From 8. | 498 | 491 |

TABLE E-1. (Continued)

| Test Number | C-319-3* | C-319-4* | C-319-5* |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 700 | 700 | 700 |
| 2. Bed temperature, °F | 1980 | 2000 | 1980 |
| 3. Starter bed | | | |
| a. Static depth, in. | 18 | 18 | 18 |
| b. Sieve size, USS | .........-8+16........ | | |
| 4. Fly ash input, lb/hr (dry) | 48 | 27 | 0 |
| a. Water content, % (a.r.) | 0 | 0 | 0 |
| b. Carbon content, % dry | 65.0 | 65.0 | -- |
| 5. Coal input, lb/hr | 6.8 | 23.2 | 42.0 |
| a. Carbon content, % | 66.0 | 66.0 | 66.0 |
| 6. Total carbon input, lb/hr | 35.6 | 32.4 | 27.7 |
| 7. Fly ash output, lb/hr | 18.8 | 15.3 | 7.6 |
| a. Carbon content, % | 15.0 | 15.2 | 17.0 |
| 8. Carbon burned, lb/hr | 32.8 | 30.1 | 26.4 |
| 9. Combustion efficiency, % | 92.0 | 92.9 | 95.5 |
| 10. Superficial velocity, fps | 13.5 | 13.5 | 13.5 |
| 11. Cooling probe position | .........Not used....... | | |
| **Flue Gas Data** | | | |
| 12. $CO_2$, % | 10.9 | 10.2 | 10.0 |
| 13. $O_2$, % | 9.5 | 9.8 | 9.8 |
| 14. CO, % | 0 | 0 | 0 |
| 15. $SO_2$, ppm | 1300 | 1600 | 2000 |
| 16. NO, ppm | 490 | 440 | 420 |
| 17. Hydrocarbons, ppm | 15 | 22 | 35 |
| 18. Particulate emission | | | |
| a. To stack, lb/hr | .47 | .19 | .17 |
| b. Carbon content, % | 7.0 | 8.3 | -- |
| **Heat Balance Data** | | | |
| 19. Losses, kBtu/hr | | | |
| a. Probe | 0 | 0 | 0 |
| b. Wall | 49 | 51 | 51 |
| c. Hood | 13 | 16 | 16 |
| d. Gas | 386 | 403 | 412 |
| e. Dust | 8 | 6 | 3 |
| f. Total | 456 | 476 | 482 |
| 20. Input, kBtu/hr From 8. | 481 | 489 | 489 |

*Heat balance data adjusted for hydrogen content of coal.

TABLE E-1. (Continued)

| Test Number | C-320-1 | C-320-2** | C-320-3*** |
|---|---|---|---|
| **Reactor Data** | | | |
| 1. Air rate, lb/hr | 620 | 620 | 620 |
| 2. Bed temperature, °F | 1980 | 1970 | 1970 |
| 3. Starter bed | | | |
| a. Static depth, in. | 20 | 20 | 20 |
| b. Sieve size, USS | -8x0 | | |
| 4. Fly ash input, lb/hr (dry)* | 167 | 174 | 208 |
| a. Carbon content, % | 27.9 | 27.5 | 19.2 |
| 5. Carbon input, lb/hr | 46.5 | 47.9 | 40.0 |
| 6. Fly ash output, lb/hr | 63 | 159 | 215 |
| a. Carbon content, % | 9.46 | 17.0 | 3.59 |
| b. Sulfur content, % | 2.49 | 2.02 | 2.66 |
| c. Carbon output, lb/hr | 6.0 | 26.9 | 7.7 |
| 7. Carbon burned, lb/hr | 40.5 | 21.0 | 32.3 |
| 8. Combustion efficiency, % | 87.1 | 43.8 | 80.7 |
| 9. Superficial velocity, fps | 13.1 | 12.7 | 12.9 |
| 10. Cooling probe position | OUT | 1 tube lowered | OUT |
| 11. Fuel heat input, kBtu/hr | 570 | 296 | 455 |
| 12. Input sulfur content, % | 2.42 | 2.42 | 2.92 |
| 13. Sulfur input, lb/hr | 4.0 | 4.2 | 6.1 |
| 14. Sulfur output, lb/hr | 1.6 | 3.2 | 5.7 |
| **Flue Gas Data** | | | |
| 15. $CO_2$, % | 15.0 | 20.0 | 16.0 |
| 16. $O_2$, % | 4.3 | 0.5 | 0.8 |
| 17. CO, % | 0 | 0 | 0 |
| 18. $SO_2$, ppm | 350 | 2150 | 4400 |
| 19. HC, ppm | 8 | 7 | 200 |
| 20. Particulate emission | | | |
| a. To stack, lb/hr | No data | | |
| b. Carbon content, % | 4.4 | 5.3 | 3.6 |
| **Heat Balance , kBtu/hr** | | | |
| 21. Probe loss | 62 | 177 | 92 |
| 22. Wall loss | 5 | 13 | 18 |
| 23. Hood loss | 91 | 31 | 68 |
| 24. Flue gas loss | 342 | 330 | 366 |
| 25. Ash loss | 26 | 66 | 89 |
| 26. Total losses | 526 | 617 | 633 |

See footnotes next page.

* Feed material C-320-1 and C-320-2: Fly ash containing $CaSO_4$ from Run FBM B-16. C-320-3: Synthetic fly ash: gypsum 10.6%, ash 26.4%, $CaCO_3$ 34%, coal 29%.

** Also at $O_2$= 1.0,$SO_2$=550; $O_2$=0.8, $SO_2$=750; $O_2$=0.4, $SO_2$=2400; $O_2$=0.1, $SO_2$=4100.

*** Also at $O_2$=1.7, $SO_2$=4300; $O_2$=1.9, $SO_2$=3800; $O_2$=2.0, $SO_2$=3400.

# APPENDIX F. STATISTICAL ANALYSIS AND MODELLING OF FBC FLY ASH CARBON-BURNUP DATA

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## PART I*

### INTRODUCTION

A series of definitive sequential experiments were carried out to generate response data at near-optimum conditions with the exception of several exploratory runs used to evaluate various alternatives such as coil cooling and bed particle size distribution.

Table F-1 includes a summary of the operating conditions observed during the series of runs and the measured response levels.

For the purposes of an experimental test program, most of the variables included in Table F-1 are well defined. However, in the case of bed temperature it is, in effect, a pseudo-independent variable. Its set point and control can be achieved by varying fly ash (at specified percent carbon levels) and air rates. In addition, maximum temperature limitations (~2000°F) can be maintained at higher carbon feed rates through the use of cooling coils. A theoretical heat balance relationship between carbon and air rates related to temperature level was developed by PER and summarized in their Progress Report Nos. 2, 3, and 4.

### REGRESSION MODELLING

The independent variables which were considered in various alternative regression modes are given in Table F-1. The principal response, CE-combustion efficiency, served as the primary criterion for evaluating the alternative models. The other response variables, included in Table F-1, were also fitted to the models but given secondary priority for selecting the final model form. Empirical regression models of the type

$$y = b_0 + b_1 x_1 + --- + b_k x_k$$

were successively fitted to the data and evaluated through their corresponding F-ratios. Here the F-ratio is a statistical measure of the variance in the response data which is explained by the model divided by the residue (unexplained) variance. If there were no relationship between the independent variables and the response, the ratio would tend to take on a value of one. In general, the higher the ratio, the more reliable the model is in explaining the trends in the data.

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[*] Editor's Note: The report is reproduced here as submitted by the author July 1970, with minor editorial corrections.

As previously discussed, bed temperature is, in effect, a pseudo-independent variable. Ordinarily, it would not be included in a model in combination with air rate, feed rate and percent carbon, since its level is primarily a function of these variables. However, with the implementation of coil cooling, alternative combinations of the causative variables, in general, can be specified for a specified temperature level. Without inclusion of the temperature variable, no satisfactory model could be developed. This is a direct consequence of the fact that no satisfactory way could be developed to quantify the alternate variable-coil cooling.

In structuring the empirical regression model, two alternate groupings of the fly ash variables were considered. These included:

GROUP 1 - FLY ASH FEED RATE
PERCENT CARBON

GROUP 2 - CARBON FEED RATE
INERT FEED RATE

The basic nature of the alternate groupings tends to take on different interpretations when they are expanded into cross-product and required terms.

Based on exploratory regression analysis, it was found that the second grouping resulted in higher F-ratios than the first group. In particular, it was found that a cross-product term (interaction) between the two variables in group 2 was highly significant and added appreciably to the reliability of the regression model. Its counterpart through the group 1 combination is not directly definable. Because of the improved reliability with the group 2 variables, only those corresponding results will be reported.

Initially, all of the experimental data, 47 test results, were included in the regression analysis. However, as anticipated, it was found that the data collected on Runs C-302 and C-303 significantly reduced the reliability of the model. The first two runs were carried out with high moisture fly ash. This moisture seems to have adversely influenced the response data and trends.

An additional point, C-314-4, was carried out, as planned, with auxiliary coal feed. There was no simple way of including this run point, and it was therefore excluded from the analysis. The remaining 38 points were used to develop the final model form.

Various alternative forms, for representing the combustion efficiency response, were evaluated. These included

CE (as reported)

$$LOG(1 - \frac{CE}{\alpha})$$

LOG(CE)

The purpose here was to develop a response form consistent with theoretical considerations. That is, a response model which could not predict above the theoretical limit of 100% combustion efficiency. Therefore, the alternative response form

$$LOG(1 - \frac{CE}{\alpha})$$

[with $\alpha$ = 95-100% corresponding to the theoretical limit of operation] would be a more satisfactory response. A summary of the respective F-ratios for CE, LOG (CE), and various values of $\underline{\alpha}$ indicate that the form (CE) results in the best compromise. This is a consequence of the fact that as $\underline{\alpha}$ gets relatively large $LOG(1 - \frac{CE}{\alpha})$ is inversely proportional to (CE)

| FORM | F-RATIO |
|---|---|
| CE | 74.5 |
| -- | -- |
| $LOG(1 - \frac{CE}{95})$ | 46.2 |
| $LOG(1 - \frac{CE}{101})$ | 69.2 |
| $LOG(1 - \frac{CE}{110})$ | 72.1 |
| $LOG(1 - \frac{CE}{120})$ | 74.2 |
| $LOG(1 - \frac{CE}{145})$ | 75.1 |
| -- | -- |
| LOG(CE) | 73.8 |

As reflected in the above table, relatively large $\alpha$ values of 120% + are required to attain the reliability level of the CE response. Therefore, this form of the response, $LOG(1 - \frac{CE}{\alpha})$, still does not satisfy the theoretical limit 95-100%, and its form is not felt to be justified.

From the standpoint of interpretation, the form CE is not too unsatisfactory, since over the range of data unrealistic response values are not defined.

The ultimate form of the model, which best related all of the data, was found to be

$$\begin{aligned} RESP = {} & b_0 + b_1(TEMP) + b_2(AIR\ RATE) \\ & + b_3(BED\ DEPTH) + b_4(CARB.\ RATE) \\ & + b_5(INERT\ RATE) + b_6(CARB.\ R)(INERT\ R) \end{aligned}$$

Over the range of the observed data, curvature terms were found to be insignificant.

A summary of the reliabilities of the models expressed in terms of their respective F-ratios and $R^2$ (the explained variation divided by the total observed variation) is given in the following table:

| RESPONSE | F-RATIO | $R^2$ |
|---|---|---|
| CE | 74.5 | .935 |
| $CO_2$ | 23.0 | .816 |
| $O_2$ | 26.3 | .836 |
| $SO_2$ | 9.2 | .642 |
| NO | 6.0 | .535 |

The computed regression coefficients for the respective models are given in Table F-2. Observed and calculated values of combustion efficiency are shown in Table F-3.

## TEMPERATURE RESPONSE

As previously discussed, bed temperature is a function of other independent variables. As derived by PER, temperature level is extremely sensitive to air and carbon rates (at a particular combustion efficiency rate). In addition, based on an analysis of the test data, it is also quite sensitive to the percent carbon in the fly ash and bed conditioning. This latter effect was observed in runs prior to C-308. For this reason, Run C-308 was carried out under varying conditionings involving initial bed temperatures prior to the run. In particular, it was found that carbon rate could be increased from 38 to 42 lb/hr (all other variables at the same level) through nominal bed conditioning.

An attempt was made to develop a temperature model in terms of the observed operating conditions (with those runs in which there was no coil cooling). However, no satisfactory model could be developed. The inherent sensitivity of bed temperature to measured and unmeasured variables (such as conditioning discussed above) is best exemplified by comparing several of the runs:

| RUN NO. | TEMP | AIR RATE | CARBON RATE | (CE) |
|---|---|---|---|---|
| C-306-1 | 2100 | 600 | 41 | (86) |
| -4 | 1800 | 600 | 43 | (73) |
| C-307-1 | 1800 | 600 | 41 | (76) |
| -3 | 2140 | 600 | 44 | (91) |

During runs C-306 and C-307, a 300°F temperature range was observed at essentially the same operating conditions.

The above data also illustrates the separate effect of temperature on combustion efficiency. The observed large differences in efficiency are totally caused by the temperature differential since all other variables are at essentially the same level (at a temperature of 2100°F, a slight increase in carbon rate, with coil cooling, would have a negligible effect on the efficiency.

## RECOMMENDATIONS AND CONCLUSIONS

Based on an analysis of the data, it is apparent that a temperature control loop is required for final modelling and optimization. It is recommended that a fly-ash feed rate controller, adapted to temperature set point (at the maximum allowable level ~2000°F), be installed with the CBC. This will be required to achieve maximum performance.

With a built-in controller and operations data over an air rate range and carbon content in fly ash, a refined model can be developed to attain the fine balance between throughput, efficiency, and oxygen level.

Optimization of the present model is not justified at the present time due to the difficulty of handling the temperature variable. As previously discussed, given the operating conditions of the CBC, the model will predict (over the range of the data) the combustion efficiency and oxygen level quite well. However, there does not seem to be any ready method for predicting bed temperature at a specified air and fly-ash rate at a particular carbon level (% carbon in the fly ash). This difficulty is a consequence of the extreme sensitivity of temperature level to operating conditions.

In general, most of the directional effects of the operating variables are consistent. For example, a deeper bed is appreciably more desirable than a shallow bed. This results in a higher combustion efficiency and lower oxygen level.

Coil cooling increases carbon throughput but because of decreased combustion efficiency, results in a negligible increase in rate of carbon burned. There is a decrease in oxygen level, but it is not sufficient to justify the decreased combustion efficiency. Therefore, it does not seem advisable to utilize coil cooling.

Two levels of average particle size were included in the analysis. These results indicate a slightly better combustion efficiency with the -16 +22 particle size.

## TABLE F-1. EXPERIMENTAL TEST CONDITIONS AND RESULTS

| RUN NUMBER | TEMP | AIR RATE | BED DEPTH | FEED RATE | % C | CARBON RATE | INERT RATE | PART SIZE | COOL-ING COIL | CE | $CO_2$ | $O_2$ | $SO_2$ | NO |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| C-302-1 | 1800 | 760 | 18 | 73 | 55 | 40 | 33 | + |  | 90 | 11.5 | 9.1 | 350 | 560 |
| -2 | 2140 | 760 | 18 | 105 | 51 | 53 | 52 | + |  | 83 | 14.2 | 6.4 | 800 | 670 |
| -3 | 1780 | 820 | 18 | 72 | 61 | 44 | 28 | + |  | 86 | 10.8 | 9.7 | 550 | 570 |
| 303-1 | 1750 | 330 | 18 | 41 | 51 | 21 | 20 | + |  | 65 | 9.8 | 10.8 | 320 | 470 |
| -2 | 1980 | 330 | 18 | 46 | 53 | 24 | 22 | + |  | 66 | 11.5 | 9.0 | 600 | 480 |
| -3 | 1750 | 500 | 18 | 68 | 57 | 39 | 29 | + |  | 65 | 12.0 | 8.5 | 320 | 410 |
| -4 | 1860 | 500 | 18 | 67 | 55 | 37 | 30 | + |  | 74 | 13.2 | 7.4 | 500 | 570 |
| -5 | 1970 | 500 | 18 | 71 | 55 | 39 | 32 | + |  | 78 | 14.5 | 6.2 | 700 | 600 |
| 306-1 | 2100 | 600 | 18 | 106 | 39 | 41 | 65 | - |  | 86 | 14.3 | 6.2 | 800 | 720 |
| -2 | 1750 | 860 | 18 | 134 | 42 | 56 | 78 | - |  | 70 | 11.0 | 9.3 | 750 | 560 |
| -3 | 2120 | 775 | 18 | 130 | 40 | 53 | 77 | - |  | 89 | 14.6 | 5.8 | 800 | 770 |
| -4 | 1800 | 600 | 18 | 105 | 41 | 43 | 62 | - |  | 73 | 12.4 | 8.2 | 750 | 520 |
| 307-1 | 1800 | 600 | 17 | 106 | 39 | 41 | 65 | + |  | 76 | 13.2 | 7.5 | 500 | 410 |
| -2 | 2100 | 330 | 18 | 65 | 39 | 26 | 39 | + |  | 90 | 16.8 | 3.5 | 600 | 500 |
| -3 | 2140 | 600 | 18 | 110 | 40 | 44 | 66 | + |  | 91 | 16.0 | 4.8 | 650 | 550 |
| -4 | 1750 | 330 | 18 | 73 | 40 | 29 | 44 | + |  | 63 | 13.4 | 6.9 | 700 | 370 |
| 308-1 | 1900 | 600 | 18 | 87 | 43 | 38 | 49 | + |  | 81.5 | 13.4 | 7.1 | 600 | 520 |
| -2 | 1900 | 600 | 18 | 100 | 42 | 42 | 58 | + |  | 81.3 | 14.5 | 6.2 | 750 | 500 |
| -3 | 1900 | 600 | 18 | 101 | 42 | 43 | 59 | + |  | 80.6 | 15.0 | 5.5 | 800 | 520 |
| 310-1 | 2040 | 700 | 18 | 105 | 44 | 46 | 59 | + |  | 84.6 | 13.2 | 7.1 | 700 | 460 |
| -2 | 2000 | 700 | 18 | 125 | 44 | 55 | 70 | + | IN | 75.8 | 13.8 | 6.4 | 720 | 420 |

BED PARTICLE SIZE CODE

| | |
|---|---|
| + | -16 +22 (fine) |
| - | -8 +16 (coarse) |

TABLE F-1. (Continued)

| RUN NUMBER | TEMP | AIR RATE | BED DEPTH | FEED RATE | % C | CARBON RATE | INERT RATE | PART SIZE | COOL-ING COIL | CE | $CO_2$ | $O_2$ | $SO_2$ | NO |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| C-311-1 | 2000 | 700 | 18 | 158 | 33 | 52 | 106 | + |  | 77.0 | 16.0 | 4.5 | 700 | 570 |
| -2 | 1930 | 700 | 18 | 190 | 33 | 63 | 127 | + | IN | 66.7 | 16.6 | 3.6 | 800 | 470 |
| -3 | 2000 | 600 | 18 | 250 | 31 | 79 | 171 | + | IN | 58.0 | 18.1 | 2.1 | 850 | 470 |
| -4 | 1980 | 600 | 18 | 158 | 32 | 51 | 107 | + |  | 75.8 | 15.9 | 4.5 | 820 | 500 |
| -5 | 1950 | 800 | 18 | 232 | 32 | 75 | 157 | + |  | 64.5 | 15.8 | 4.7 | (820) | 510 |
| -6 | 1800 | 800 | 18 | 302 | 32 | 96 | 206 | + | IN | 56.2 | 16.4 | 3.9 | (820) | 350 |
| 312-1 | 2000 | 700 | 18 | 140 | 33 | 47 | 93 | + |  | 84.2 | 13.8 | 7.0 | 900 | 500 |
| -2 | 1990 | 700 | 14 | 138 | 33 | 46 | 92 | + |  | 83.8 | 13.5 | 7.2 | 900 | 480 |
| -3 | 1980 | 700 | 10 | 156 | 33 | 51 | 105 | + |  | 76.0 | 13.2 | 7.5 | 900 | 480 |
| 313-1 | 1980 | 600 | 18 | 63 | 58 | 37 | 26 | + |  | 91.0 | 13.3 | 7.5 | 1160 | 630 |
| -2 | 1980 | 700 | 18 | 74 | 58 | 43 | 31 | + |  | 87.2 | 12.9 | 8.0 | 1160 | 610 |
| -3 | 1980 | 800 | 18 | 84 | 57 | 48 | 36 | + |  | 86.0 | 12.8 | 8.1 | 1160 | 600 |
| 314-1 | 1980 | 600 | 10 | 70 | 56 | 39 | 31 | + |  | 84.8 | 13.4 | 7.4 | 1050 | 560 |
| -2 | 1960 | 700 | 10 | 82 | 56 | 46 | 36 | + |  | 82.0 | 13.0 | 7.6 | 1050 | 560 |
| -3 | 1980 | 800 | 10 | 102 | 56 | 57 | 45 | + |  | 76.0 | 13.0 | 7.8 | 1050 | 560 |
| 315-1 | 2010 | 600 | 18 | 81 | 57 | 46 | 35 | + | IN | 83.3 | 15.4 | 5.5 | 1250 | 560 |
| -2 | 2000 | 700 | 18 | 86 | 57 | 49 | 37 | + | IN | 84.5 | 15.0 | 5.8 | 1250 | 560 |
| -3 | 1980 | 800 | 18 | 92 | 56 | 52 | 40 | + | IN | 85.2 | 14.6 | 6.0 | 1250 | 540 |
| 316-1 | 2000 | 600 | 10 | 86 | 58 | 50 | 36 | + | IN | 74.5 | 14.8 | 5.8 | 1250 | 520 |
| -2 | 1990 | 700 | 10 | 94 | 58 | 58* | 36* | + | IN | 76.5 | 14.3 | 6.0 | 1300 | 510 |
| -3 | 1980 | 800 | 10 | 109 | 58 | 63 | 46 | + | IN | 72.0 | 13.8 | 6.4 | 1250 | 520 |
| 317-1 | 1980 | 700 | 22 | 95 | 54 | 52 | 43 | - |  | 82.2 | 15.5 | 4.8 | 1200 | 570 |
| -2 | 1980 | 700 | 18 | 94 | 54 | 51 | 43 | - |  | 81.5 | 15.3 | 5.3 | 1275 | 570 |
| 318-1 | 1900 | 700 | 22 | 104 | 59 | 61 | 43 | + |  | 72.5 | 15.8 | 5.0 | 1050 | 550 |
| -2 | 1950 | 700 | 18 | 97 | 58 | 56 | 41 | + |  | 79.8 | 16.0 | 4.5 | 1050 | 550 |

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*Editor's Note: These data in error. Use data from Appendix E.

### TABLE F-2. REGRESSION COEFFICIENTS

| **TERM** | CE | $CO_2$ | $O_2$ | $SO_2$ | NO |
|---|---|---|---|---|---|
| CONSTANT | -13.78 | -2.498 | 22.91 | -44.12 | -464.3 |
| TEMP | .05193 | .007547 | -.007353 | .4095 | .4257 |
| AIR RATE | .04620 | -.01190 | .01300 | .1199 | .3825 |
| BED DEPTH | .3831 | .1382 | -.1390 | -8.986 | 4.468 |
| CARB. R | -.8737 | .1589 | -.1769 | 12.60 | -2.717 |
| INERT R | -.1905 | .01259 | -.01756 | -7.608 | -.9104 |
| (CR)(IR) | .002700 | -.0002862 | .0003587 | .03046 | .007263 |

## TABLE F-3. OBSERVED AND CALCULATED COMBUSTION EFFICIENCY

| RUN | OBS | CALC | RES |
|---|---|---|---|
| C-306-1 | 86 | 88.9 | -2.9 |
| -2 | 70 | 71.7 | -1.7 |
| -3 | 89 | 89.1 | -.1 |
| -4 | 73 | 72.1 | .9 |
| 307-1 | 76 | 73.3 | 2.7 |
| -2 | 90 | 90.0 | 0 |
| -3 | 91 | 88.8 | 2.2 |
| -4 | 63 | 69.0 | -6.0 |
| 308-1 | 81.5 | 82.0 | -.5 |
| -2 | 81.3 | 78.3 | 3.0 |
| -3 | 80.6 | 77.5 | 3.1 |
| 310-1 | 84.6 | 87.3 | -2.7 |
| -2 | 75.8 | 78.3 | -2.5 |
| 311-1 | 77.0 | 78.6 | -1.6 |
| -2 | 66.7 | 68.0 | -1.3 |
| -3 | 58.0 | 59.5 | -1.5 |
| -4 | 75.8 | 73.4 | 2.4 |
| -5 | 64.5 | 67.7 | -3.2 |
| -6 | 56.2 | 53.8 | 2.4 |
| C-312-1 | 84.2 | 82.3 | 1.9 |
| -2 | 83.8 | 81.0 | 2.8 |
| -3 | 76.0 | 75.1 | .9 |
| 313-1 | 91.0 | 89.0 | 2.0 |
| -2 | 87.2 | 88.4 | -1.2 |
| -3 | 86.0 | 88.8 | -2.8 |
| 314-1 | 84.8 | 83.9 | .9 |
| -2 | 82.0 | 81.6 | .4 |
| -3 | 76.0 | 78.4 | -2.4 |
| 315-1 | 83.3 | 82.7 | .6 |
| -2 | 84.5 | 84.4 | .1 |
| -3 | 85.2 | 85.5 | -.3 |
| 316-1 | 74.5 | 76.0 | -1.5 |
| -2 | 76.5 | 73.8* | 2.7* |
| -3 | 72.0 | 73.8 | -1.8 |
| 317-1 | 82.2 | 82.2 | 0 |
| -2 | 81.5 | 81.4 | .1 |
| 318-1 | 72.5 | 71.2 | 1.3 |
| -2 | 79.8 | 76.2 | 3.6 |

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*Editor's Note: These results in error; read 76.5 for CALC and 0 for RES.

## PART II.[1] REGRESSION FOR PARTICULATE EMISSION

A regression model was developed using particulate emission data furnished by Pope, Evans and Robbins. The same run data, previously reported Runs 306 through 319, were used for this analysis.

Several alternative forms for the response and different combinations of the independent variables were tried. These alternatives led to the following results in terms of an F-ratio criteria.[2]

| MODEL | RESPONSE | F-RATIO |
|---|---|---|
| Type 1 | P.E. | 28.0 |
| Type 1 | Log (P.E.) | 18.3 |
| Type 2 | P.E. | 20.5 |
| Type 3 | P.E. | 16.5 |

### TYPE 1

$b_o + b_1$ (Bed Temperature - °F)

$+ b_2$ (Air Rate - lbs/hr

$+ b_3$ (Static Bed Depth, incites)

$+ b_4$ (Carbon Feed Rate, lbs of C/hr

$+ b_5$ (Inert Feed Rate, lbs of I/hr

$+ b_6$ (C X I)

### TYPE 2

Same as Type 1 but excluding $b_1$, $b_2$, and $b_3$.

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[1] Editor's Note: The report is reproduced here as submitted by the author in January 1972, with minor editorial corrections.

[2] The F-ratio is the explained variation or variance (as explained by the associated assumed model) divided by the unexplained variation.

## TYPE 3

Same as Type 1 but with Log (Carbon F.R.) and Log (Inert F.R.) and excluding $b_6$.

On the basis of these calculations it may be concluded that both carbon and inert feed rates and their interaction are extremely important variables as they affect particulate emissions (as well as the other responses that were previously reported.

The regression coefficients for Type 1 model with the P.E. response are:

$b_o = 15.57$

$b_1 = .006639$

$b_2 = .001571$

$b_3 = .02362$

$b_4 = .03528$

$b_5 = .005357$

$b_6 = .0003935$

The residual standard deviation for this model is s = 0.75. Table F-4 includes a summary of the reported P.E. determinations and the corresponding model predictions.

The model is quite adequate but could be improved without too much additional work if it were required. This would necessitate a more careful screening of the reported P.E. data to screen out gross errors. It was reported that the P.E. results are not highly reliable and this is reflected by the relatively large residual variance s = .75. Therefore, additional refinement of the model is not sufficient by itself, but rather a combination of model development and data screening would be required to improve the overall reliability. This would only be justified if an improved model were needed.

## TABLE F-4. OBSERVED AND CALCULATED PARTICULATE EMISSIONS

| RUN NO. | REPORTED | CALCULATED | RUN NO. | REPORTED | CALCULATED |
|---|---|---|---|---|---|
| 306-1 | 1.27 | 1.06 | 312-1 | 2.2 | 2.18 |
| 306-2 | 4.15 | 3.19 | 312-2 | 2.3 | 2.13 |
| 306-3 | 1.71 | .85 | 312-3 | 1.5 | 2.43 |
| 306-4 | 2.97 | 2.97 | 313-1 | 1.4 | 1.12 |
| 307-1 | 3.0 | 3.05 | 313-2 | 1.0 | .93 |
| 307-2 | .4 | 1.23 | 313-3 | 1.7 | .77 |
| 307-3 | 1.5 | .79 | 314-1 | .9 | .99 |
| 307-4 | 4.1 | 3.57 | 314-2 | .5 | .92 |
| 308-1 | 2.7 | 2.09 | 314-3 | .6 | .65 |
| 308-2 | 2.5 | 2.23 | 315-1 | .7 | .91 |
| 308-3 | 3.1 | 2.22 | 315-2 | .7 | .80 |
| 310-1 | .88 | 1.11 | 315-3 | .8 | .79 |
| 310-2 | 1.1 | 1.57 | 316-1 | .9 | .72 |
| 311-1 | 1.32 | 2.52 | 316-2 | .8 | .66 |
| 311-2 | 2.78 | 3.69 | 316-3 | .9 | .57 |
| 311-3 | 7.50 | 5.22 | 317-1 | .68 | 1.12 |
| 311-4 | 1.77 | 2.83 | 317-2 | .6 | 1.05 |
| 311-5 | 5.3 | 4.62 | 318-1 | .72 | 1.49 |
| 311-6 | 7.1 | 8.29 | 318-2 | .75 | 1.10 |
|  |  |  | 319-1 | .60 | .88 |
|  |  |  | 319-2 | .61 | .70 |

# APPENDIX G. GAS TEMPERATURE GRADIENT IN A HOT FLUIDIZED BED

The problem of measuring the gas temperature gradient at the base of a hot fluidized bed is complicated by radiation heat gain from bed particles in the vicinity of the thermocouple. The problem is analagous to that of measuring temperature in very hot gas streams in a non-adiabatic duct, except here heat is lost by radiation from the thermocouple. In the latter instance, the usual practice is to reduce the radiation loss by shielding and increase the convective heat gain by high gas velocities. In the fluidized bed operating at above 1600°F, it is not possible to shield out the radiation field effectively without disturbing the factors which influence the local temperature.

An estimate was made of the gas temperature gradient as it enters the bed, based on convective exchange of heat between the gas and the bed particles. It was assumed in the analysis that the gas moved in plug flow through a back mix reactor of particles and that the particle temperature remained constant at the total bed temperature. This latter assumption seems valid in view of the fact that no temperature gradients are found when burning coal.

The following function was determined for the variation of gas temperature with distance:

$$T_g = T_b - (T_b - T_o)e^{-kx}, \quad k = \frac{hA_i}{GC_pL} \tag{G-1}$$

where $T_g$ and $T_b$ are the gas and bed temperatures respectively, h is the particle-to-gas heat transfer coefficient, $A_t/L$ the particle surface per unit bed depth, G the mass gas flow, $C_p$ the gas heat capacity, and x the distance above the grid, $T_o$ is the initial gas temperature.

The function is plotted in Figure G-1 for a 16-mesh bed particle size at a 1600°F bed temperature. The particle surface was estimated from that of close packed spheres with a 50% bed expansion. The mass flow was taken at 780 lb/hr and the particle-to-gas transfer coefficient was estimated from the relation*:

$$\frac{hd}{k} = 0.016\left(\frac{DG}{\mu}\right)^{1.3}\left(\frac{C_p\mu}{k}\right)^{0.67} \qquad \text{(G-2)}$$

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*Frantz, J. F., Chem. Eng. Prog. 57, 35(1961)

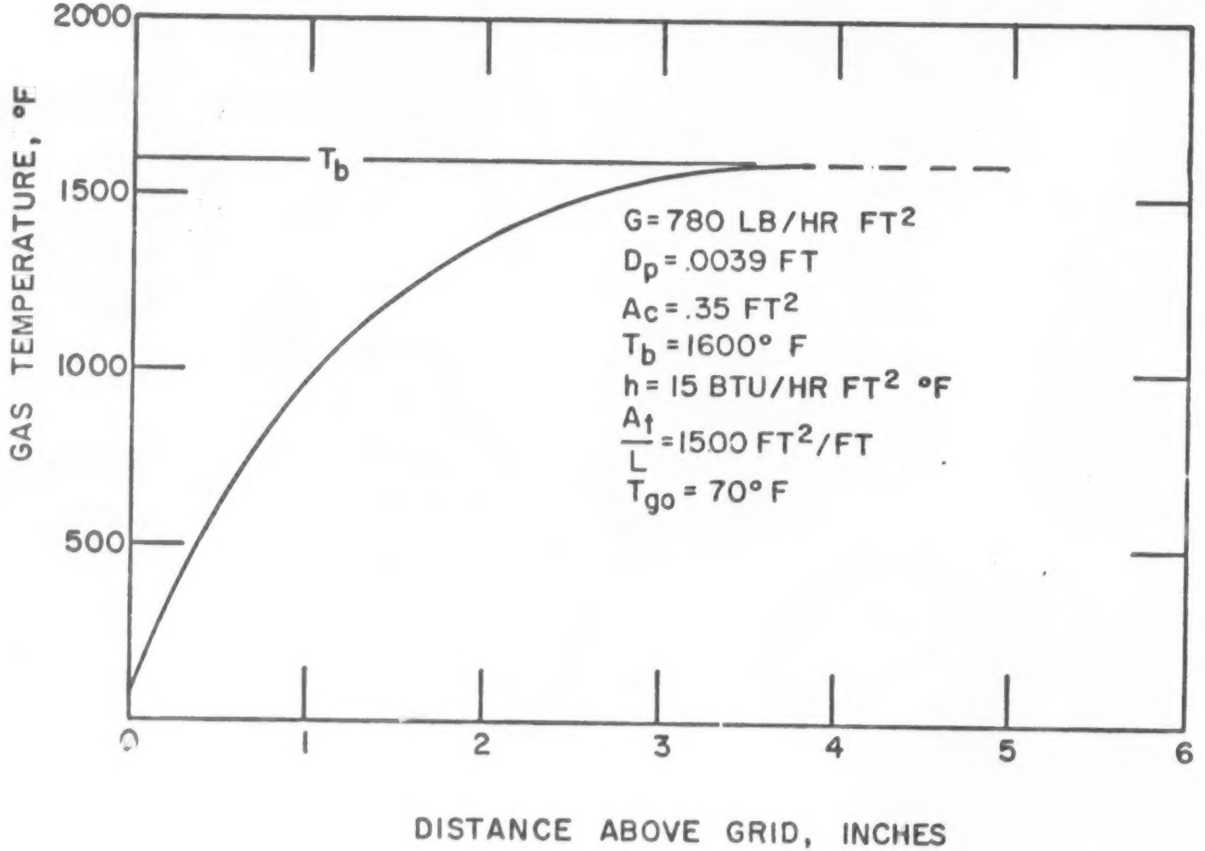

FIGURE G-1 ESTIMATED TEMPERATURE GRADIENT AT THE BASE OF THE FLUIDIZED-BED

The temperature gradient or the rate of rise with distance at x = 0 is:

$$\frac{dT_g}{dx} = \frac{hA_t}{GC_pL}(T_b - T_o) \qquad \text{(G-3)}$$

The gas temperature variation with time is given by the function:

$$T_g = \frac{T_o T_b}{T_o + (T_b - T_o)e^{-zt}} \quad , z = \frac{T_b h A_t}{k C_p A_c L} \qquad \text{(G-4)}$$

Where $A_c$ is the effective interparticle cross section area in the bed (for flow) and k a constant equal to 39.6 lb °R/ft$^3$.

In this case, the temperatures are expressed in degrees Rankine. Derivations of these relations are on file.

The temperature-time relation is plotted in Figure G-2 using an estimated value of .35 ft$^2$ for $A_c$. The curves presented in the two figures show that the gas temperature reaches the bed temperature in about 10-12 milliseconds and in a distance of .3 inches at the stated mass flow. It is interesting to note that the rise time is independent of mass flow whereas the rise distance is proportional to mass flow. The distance-time function is also shown. This curve expresses the function:

$$X_t = \frac{GT_b}{kA_c}\left[t - \frac{1}{z}\log\frac{T_b}{T_o + (T_b - T_o)e^{-zt}}\right] \qquad \text{(G-5)}$$

These results indicate that fine fly ash particles, whose temperature should follow that of the gas, reach combustion temperature in a few milliseconds. The distance travelled before rapid combustion begins is on the order of three tenths of an inch.

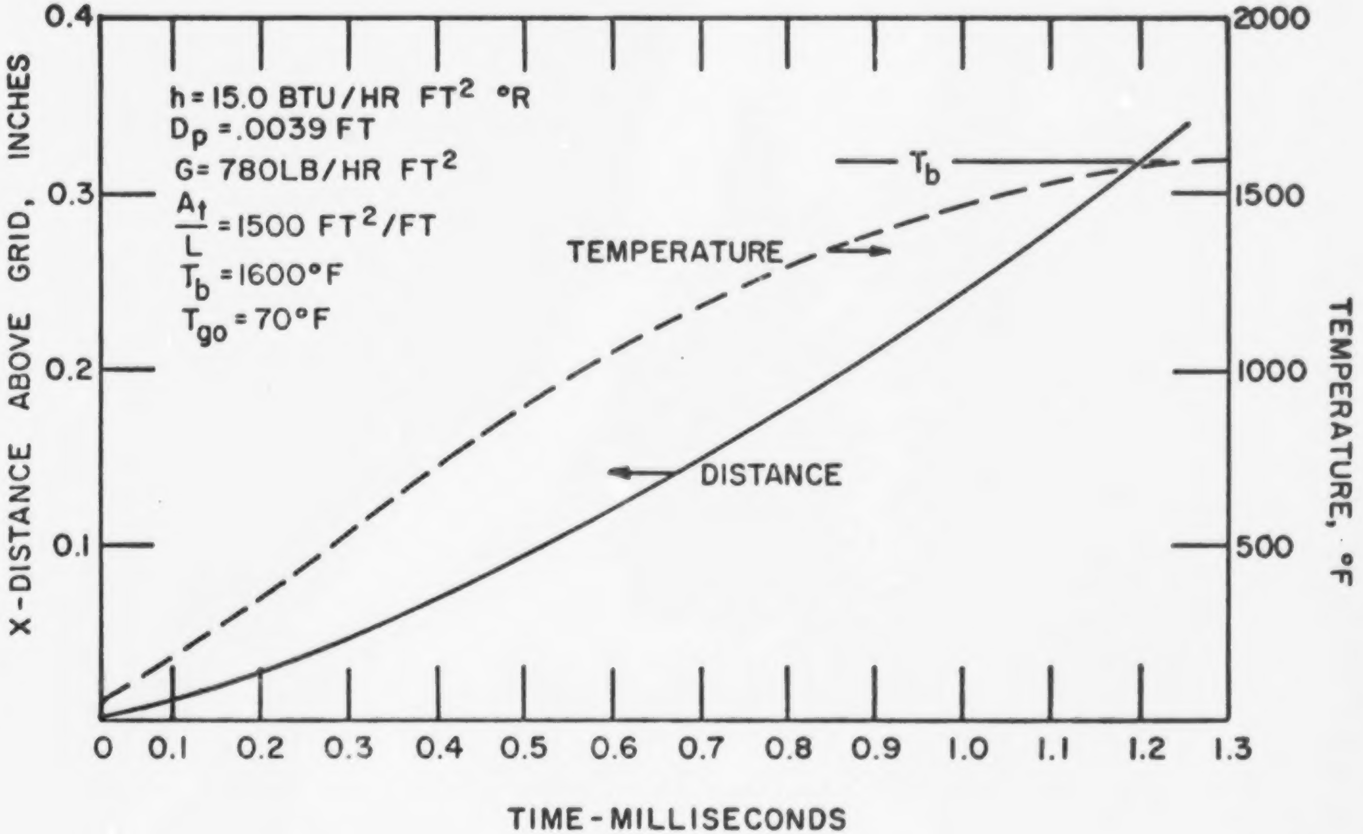

FIGURE G-2 TIME VARIATION OF GAS TEMPERATURE AND DISTANCE TRAVELED THROUGH FLUIDIZED-BED

## APPENDIX H. A THEORETICAL ANALYSIS OF SULFUR RETENTION PHENOMENA (THE PSEUDOSULFATE THEORY)

A reexamination of previous $SO_2$ control results was made. A theoretical explanation of the performance of a fluidized-bed boiler in once-through and regenerative modes of lime desulfurization is described.

The factors that limit the limestone removal of $SO_2$ from combustion gases in a coal-fired boiler may be understood in the context of the following considerations. Assume that the following equilibria and rates are involved:

$$\mathrm{CaCO_3} \underset{k_1'}{\overset{k_1}{\rightleftharpoons}} \mathrm{CaO} + \mathrm{CO_2} , \qquad K_1 \qquad \text{(H-1)}$$

$$\mathrm{CaO} + \mathrm{SO_2} + \frac{1}{2}\mathrm{O_2} \underset{k_2'}{\overset{k_2}{\rightleftharpoons}} \mathrm{CaSO_4} , \qquad K_2 \qquad \text{(H-2)}$$

$$\mathrm{CaCO_3} + \mathrm{SO_2} + \frac{1}{2}\mathrm{O_2} \rightleftharpoons \mathrm{CaSO_4} + \mathrm{CO_2}, K_3 = K_1K_2 \qquad \text{(H-3)}$$

The K's are the equilibrium constants, and the k's are the rate constants.

The concentration of $SO_2$ in equilibrium with solid $CaCO_3$ and $CaSO_4$, in the presence of known concentrations of $CO_2$ and $O_2$, may be calculated from the published equilibrium constant for reaction (H-3). Thus:

$$[\mathrm{SO_2}]_{\mathrm{eq.}} \text{ in ppm} = \frac{[\mathrm{CO_2}]}{K_3\ [\mathrm{O_2}]^{\frac{1}{2}}} \times 10^6 \qquad \text{(H-4)}$$

where $[SO_2]$ is in ppm (by volume), $[CO_2]$ and $[O_2]$ are the partial pressures of these gases in atm, and $K_3$ is the equilibrium constant in $1/\sqrt{\mathrm{atm}}$. The results of such calculations for concentrations of $CO_2$ and $O_2$ comparable to those present in the fluidized-bed systems are shown in curves A of Fig. H-1.

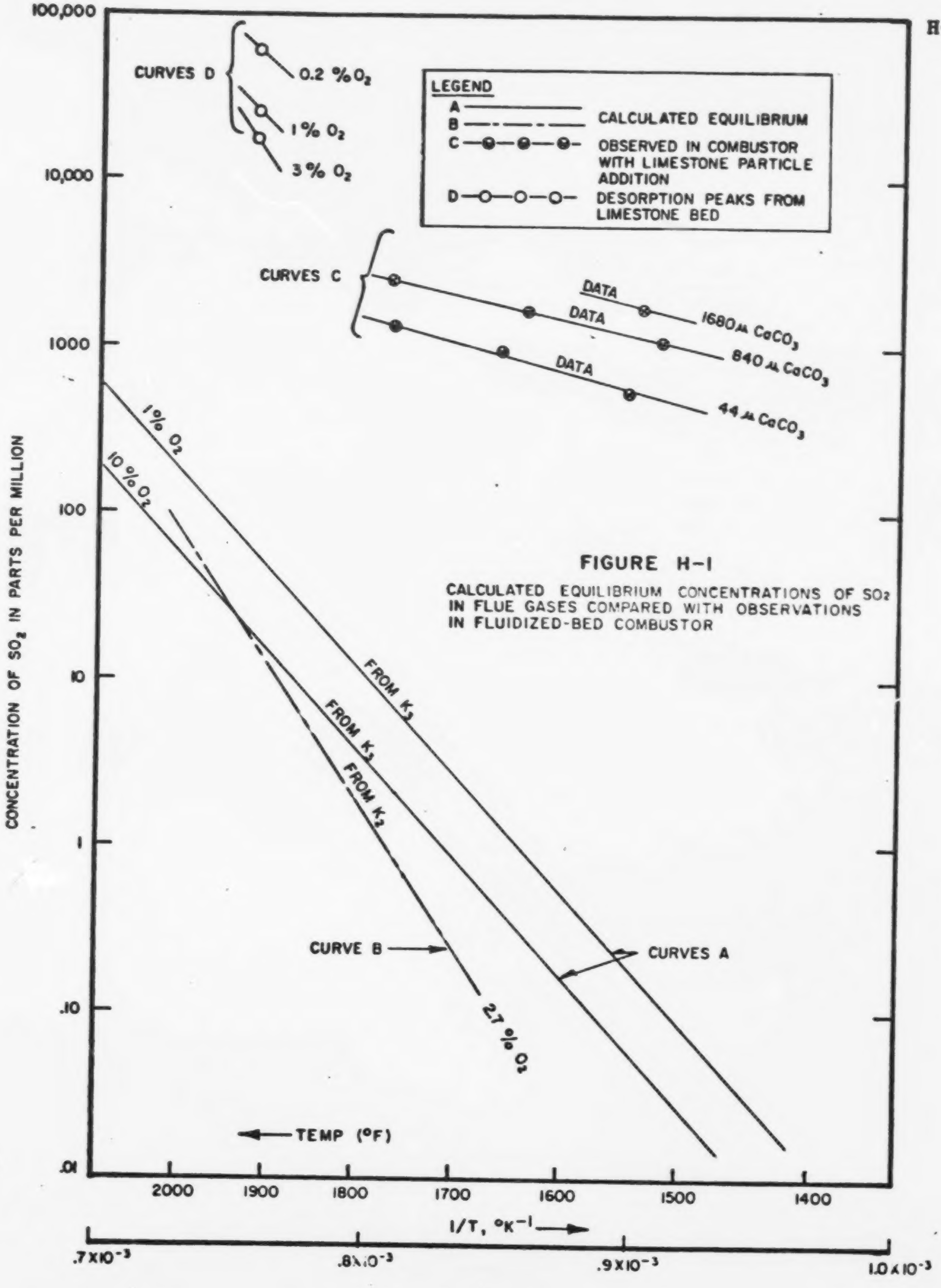

FIGURE H-1

CALCULATED EQUILIBRIUM CONCENTRATIONS OF $SO_2$ IN FLUE GASES COMPARED WITH OBSERVATIONS IN FLUIDIZED-BED COMBUSTOR

The partial pressure of $CO_2$ is taken as 0.15 atm, and two concentrations of excess oxygen are shown, 1% and 10%. The equilibrium constant is taken from the tabulation prepared for NAPCA by the FMC Corporation, "Applicability of Inorganic Solids... etc." PB 184 751. These calculations are in fair agreement with the calculations of W.T. Reid (Paper No. 69-Pwr-5, ASME - IEEE Power Generation Conference Sept. 1969), which are shown in curve B. Reid's curve is calculated from $K_2$ only, with an assumed concentration of 2.7% $O_2$. Since there is fair agreement, the equilibrium condition does not appear to be too sensitive to the presence of reaction H-1 at least at the higher temperatures.

The measured $SO_2$ concentrations from the PER experiments in coal-fired, fluidized-bed boilers are shown in the curves marked C taken from Reference 1. The data are for stoichiometric ratios Ca/S of 2.5 - 2.7 and for decreasing particle sizes.

The comparison indicates that the observed $SO_2$ concentrations (Curves C) are from two to three orders of magnitude higher than the assumed equilibrium constraints. The system is closer to equilibrium at the higher temperatures than at the lower temperatures. It is also closer to equilibrium for the smaller size additives than for the larger diameter additives.

The departure from the indicated equilibrium could be caused by one or a combination of the following factors:

(i) The process is rate-limited by a slow readsorption-recombination rate constant, $k_2$, for the formation of the sulfate.

(ii) The system is controlled by equilibria other than those indicated.

(iii) Non-isothermal temperature distributions between the solid surfaces of the bed and the gas phase.

The observations generally indicate that the first factor, (i), the rate limitation for the forward reaction rate-constant, $k_2$, may be the reason for the observed reductions in $SO_2$ being well below those expected on the basis of thermochemical equilibrium. The data are closer to equilibrium from the smaller particle size additives, which have a larger area and hence a higher $k_2$. The data are closer to equilibrium for higher Ca/S ratios, which also have larger sorbent areas. These two effects are to be expected for a rate-limiting, heterogeneous reaction in which $k_2$ is expected to be directly proportional to some power of the surface area of solid reactant. In fact, the data presented in Reference 1, can be used to infer that $k_2$ varies as $d^{\frac{1}{2}}$ or $d^{\frac{1}{3}}$ where d is the particle diameter. This indicates that $k_2$ is controlled only mildly by the adsorption rate of $O_2$ and $SO_2$, and that major control is exerted by diffusion processes on the surface or interior to the solid. This viewpoint is consistent with the observation that limestone, $CaCO_3$, is more effective than pre-calcined lime, CaO. The carbonate decomposes readily, and the desorbing $CO_2$ creates fresh, internal pathways, and a large surface area, that are readily available to the adsorbing $SO_2$ and $O_2$.

It is possible that factor (ii) is a contributing cause. The following additional equilibria may influence the system:

$$Ca(OH)_2 + SO_2 + \frac{1}{2} O_2 \rightleftarrows CaSO_4 + H_2O \qquad \text{(H-5)}$$

$$CaO + SiO_2 \rightleftarrows CaSiO_3 \qquad \text{(H-6)}$$

$$CaSiO_3 + SO_2 + \frac{1}{2} O_2 \rightleftarrows \underline{SiO_2} + CaSO_4 \qquad \text{(H-7)}$$

Reaction H-5 is probably not too significant above 800°F according to Reid, and reactions H-6 and H-7 would probably not be controlling in a steady-state system where fresh limestone is being added continuously, since it involves the slow solid-solid phase slagging reaction H-6. Reactions H-6 and H-7 may however become important in a static, capture-regenerative system that is cycled at high temperatures where the formation of a slag on the fresh lime surface would tend to poison the absorbent.

Factor (iii) is a real effect, but is probably not too significant in magnitude.

The data obtained for combustion with a limestone bed, (1), which naturally contains an enormous excess of limestone, is consistent with this general interpretation. The data show a wide scatter in the measured $SO_2$ concentration for given conditions of temperature and excess oxygen. In the low temperature range, 1450-1700°F, referred to as the sorption cycle, the measured $SO_2$ concentrations are below the curves C (for the much lower Ca/S ratios in the particle addition); however, they are generally still above the equilibrium curves. In the higher temperature ranges 1900-2000°F, referred to as the desorption cycle, the peak $SO_2$ concentrations are indicated as the curves D in Fig.H-1. These concentrations are substantially larger than those that would be expected for the limestone particle additives of curves C.

Nevertheless, the data can be understood in terms of factor (i), and particularly in terms of the kinetic mechanism for reaction H-1, the sorption-desorption reaction. As indicated earlier, the limestone decomposition by reaction H-1 is rapid, and the desorbing $CO_2$ from the interior of the solid creates fresh pathways that are available for the absorbing $SO_2$ and $O_2$. The kinetic mechanism of the absorption process is probably a two-step process; the first process, α, involving the chemisorption of $SO_2$ and $O_2$ on the surface of the freshly calcined limestone; the second process, β, involving the conversion of the chemisorbed species to the stable sulfate. Thus $k_2$ is a composite rate-constant for the two sequential reactions:

$$CaO + SO_2 + \frac{1}{2} O_2 \overset{\alpha}{\rightleftarrows} CaO\cdot[SO_2\cdot O]_{ads} \overset{\beta}{\rightleftarrows} CaSO_4$$

Furthermore, process β is itself rather complex, involving first the mobility of $SO_2$ and O along the surface to active sites where they react with each other and with the $Ca^{++}O^{=}$ lattice to form an initial sulfate film. Secondly, once the initial film is formed, the formation of more sulfate is limited by the interdiffusion of reactants through the existing sulfate film. [$Ca^{++}$ ions diffuse from the interior and/or $SO_2\cdot O$ diffuse from the surface.] Thus the overall rate constant, $k_2$, is probably limited by the process β which is much slower than α. This viewpoint is essentially equivalent to the prevailing view that sorbent utilization is limited by the formation of a sulfate shell on the sorbent particles.

However, this overall viewpoint also implies that the reactive lime surface is both a chemisorbing surface and a reactant. Thus, in a limestone bed containing a large surface area, a significant fraction of the $SO_2$ captured could be in the form of a chemisorbed intermediate $CaO\cdot[SO_2\cdot O]_{ads}$, or in the form of a pseudosulfate intermediate $CaSO_4\cdot[Ca^{++}O^{=}\cdot SO_3]$. It is plausible to postulate that it is these intermediates which decompose readily during the desorption cycles at 1900-2000°F, generating much higher $SO_2$ concentrations than would be expected from the stable sulfate. The TVA report of August 1969 indicates that the measured decomposition of the true sulfate $CaSO_4$ is small even at 2200°F, as the equilibrium curves predict. Thus the behavior of the limestone-bed system is also explainable in terms of the process being limited by the rate constant $k_2$. And, in particular, the behavior of the system in the sorption-desorption cycles is consistent with the existence of a chemisorbed or pseudosulfate intermediate which is less stable than the true sulfate and which undergoes its sorption-desorption equilibrium in a lower temperature range than the stable sulfate.

The observed dependence of $SO_2$ concentrations in the flue gas, on the excess $O_2$ in the system, is as would be predicted by the chemisorption rate process $\alpha$.

The difficulty of obtaining steady-state equilibrium in the experiments with a limestone bed is associated with the large mass of sorbent used relative to the sulfur present in the combusting coal. With its large surface area that is reactive to the combustion products, the limestone bed exerts a stronger influence on flue gas composition than the limestone particulates. The limestone bed is no longer the inert thermal matrix of the conventional bed material, but is rather an active reactant. If combustion conditions are altered, it takes longer for the system to reach steady-state due to the sluggishness of process $\beta$, and the large mass of reactant.

To test the possibility that pseudosulfate formation in boilers would explain the discrepancy observed by other investigators between:

1. Equilibrium $SO_2$ partial pressure at a given temperature over calcium sulfate by thermodynamic calculations, and

2. $SO_2$ descrptions actually observed in sulfated limestone regeneration under low oxygen conditions,

a test was performed.

The approach taken was to save all fly ash from an FBM run, which was mixed and analyzed: 27% carbon, 2.42% sulfur, 19.4% Ca. A separate batch of "synthetic fly ash" was then prepared as follows: inert fine ash 26.4%, limestone 34%, gypsum 10.6%, coal fines 29%. A test was then made in the FBC at 1970°F to compare $SO_2$ desorption from sulfated fly ash with that from the synthetic mixture where Ca and S are present at the appropriate ratio. Note that the "synthetic mix" sulfur content is 1.97% as $CaSO_4 \cdot 2H_2O$ plus 0.95% in the coal fines. $CO_2$ and $H_2O$ are generated by decomposition of the limestone and gypsum. The coal fines equivalent $SO_2$ in the flue gas from burning the synthetic mix is 2900 ppm. The excess above this value represents gypsum decomposition.

The FBC unit was successfully operated on fly ash containing the low value of 19% to 27% carbon. At this high dust loading there is, from the material balances, some holding of fine material in the bed. At about 2000°F and low $O_2$ levels in the flue gas, some of the $SO_2$ captured by limestone in the FBM desorbed. The $SO_2$ equilibrium ranges from 4100 ppm at 0.1% $O_2$ to 350 ppm at 4.3% $O_2$ . Most of the points shown in Figure H-2 represent flash readings without separate material balances, necessitated by the low supply of fly ash. Substitution of the synthetic mix at four levels of $O_2$ leads to $SO_2$ values of 3400 to 4400 depending on $O_2$ level; all of the $SO_2$ levels, after correction for the coal S content, are at or above the corresponding levels previously obtained with real sulfated fly ash. It appears unlikely that calcining gypsum could lead to "pseudosulfate" formation.

It would be desirable to repeat this experiment with "synthetic mix" containing coal fines of low S content, such as anthracite fines, to minimize the needed correction. Fly ash of lower Ca and higher C content would also be desirable to reduce the dust loading. Nevertheless the results indicate the "pseudosulfate" hypothesis is unnecessary to explain the observations. Also, the existing high temperature $CaSO_4$ thermodynamic data, on which the equilibrium calculations are based, need to be redetermined.

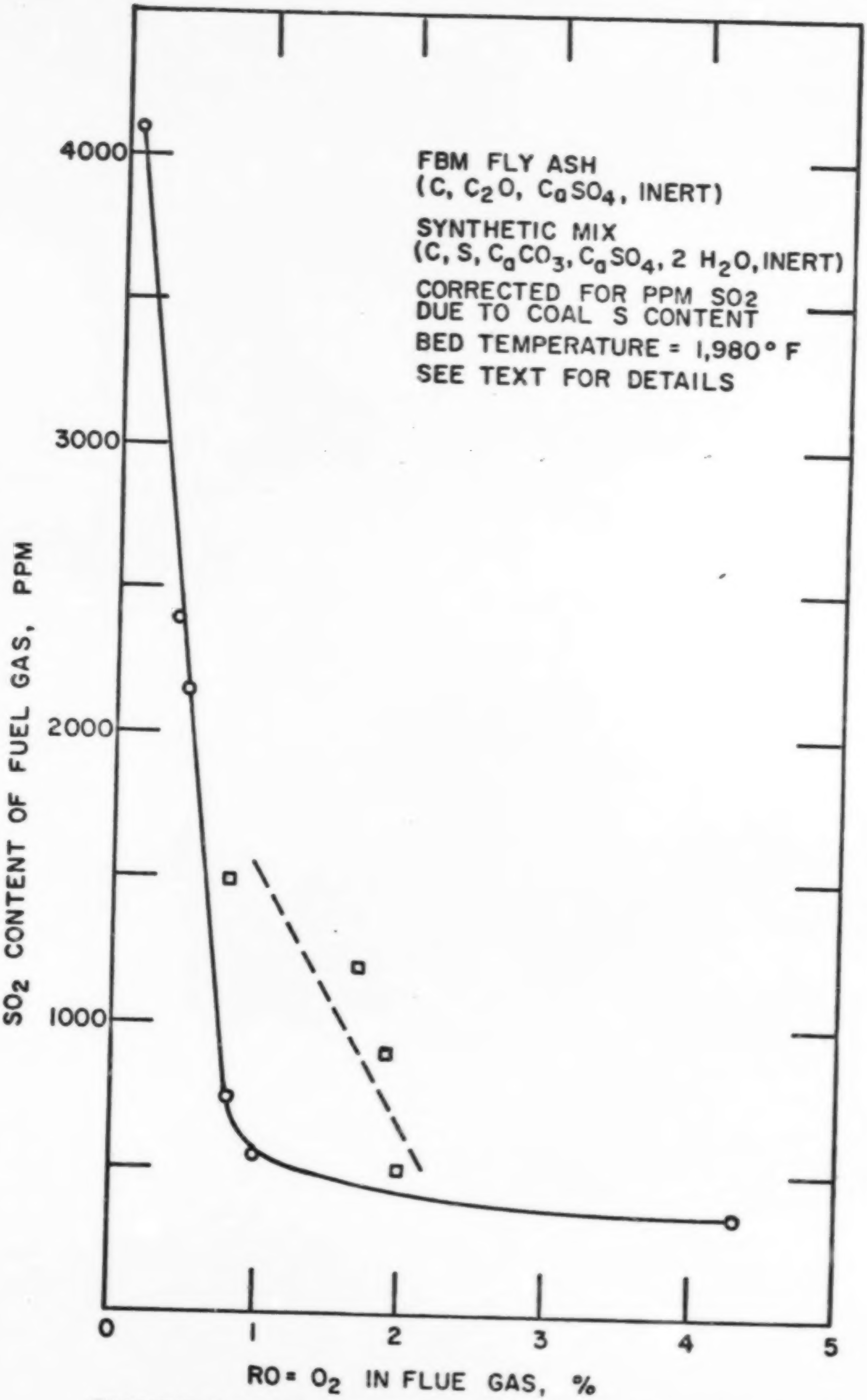

FIGURE H-2 $SO_2$ DESORPTION IN FBM FROM SULFATED FLY ASH AND SYNTHETIC MIX

## APPENDIX I. ADJUSTMENT PROCEDURE FOR CARBON-BURNUP CELL PERFORMANCE MODEL WITH COAL IN FUEL

The Carbon-Burnup Cell Performance model (Equation 1) based on tests in which carbon-bearing fly ash was the sole fuel will predict performance when coal is part or all of the fuel if the following adjustments are made:

1. The carbon content of the coal fed was increased by treating two moles of hydrogen as if they were one mole of carbon. This is based on the oxygen requirement of the two reactants for complete combustion.

   The adjusted carbon rate, $C_1$, is given by:

$$C_1 = \frac{\left[\text{Fly Ash Rate} \times \text{Wt. \%, Carbon} + \text{Coal Rate}\left\{\text{Wt. \%, Carbon} + \frac{\text{Wt. \%, H}}{2}\left(\frac{12}{2}\right)\right\}\right]}{100} \qquad (\text{I-1})$$

2. The observed combustion efficiency is adjusted to account for the total oxidation of hydrogen under the conditions of a Carbon-Burnup Cell.* The adjusted value, CE' is given by:

$$CE' = \frac{(C_i - C_o) \times 14{,}100 + C_c(\text{Wt.}\ \%, H) \times 51{,}600}{(C_i) \times 14{,}100 + C_c(\text{Wt.}\ \%, H) \times 51{,}600} \qquad \text{(I-2)}$$

where

$C_i$ = Fly ash rate x (wt %, Carbon)
+ Coal rate x (wt %, Carbon)
$C_o$ = Ash collected in cyclone x (wt %, Carbon)
$C_c$ = Coal rate

All mass flows are in pounds per hour. Note also that CE' CE because the adjustment is small and is added both above and below the line.

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* The hydrogen content of the gas is not actually measured. It is assumed that under strongly oxidized conditions none would exist. This assumption is supported by the low hydrocarbon values in the FBC exhaust.

# APPENDIX J. BED MATERIAL FLOW RATE BETWEEN REGIONS

When a fluidized-bed boiler is designed in such a way that various regions of the boiler are to operate at different temperatures, the interchange of bed material becomes a design consideration. For example, the flow of relatively cool particles into a hotter region acts as a heat sink. To compensate for the "cool-particle heat sink," either more fuel must be burned or less energy extracted by heat exchange surfaces, excess air, etc.

The flow between adjacent regions separated by a "semi-permeable" barrier may be the result of diffusion or a pumping action resulting from pressure differences. Whatever the motive power, the interchange rate should be proportional to the area of the openings between regions.

To establish the interchange rate, two tests were conducted in the FBM in which the boiler was fired with coal while the CBC remained cold and unfluidized. When stable boiler operation was achieved the burnup cell was fluidized, without fuel being added, and the CBC bed temperature rise followed. Both the transient rise and steady-state values were noted, as both might be instructive. It was found that the temperature of the bed rose more rapidly than that of the thin chamber walls. This meant that the equations describing the heat-up rate could not be integrated.

The transient heating rate is described by

$$M_c c_{p_c} \frac{dT_c}{dt} + M_b c_{p_b} \frac{dT_b}{dt} = \tag{J-1}$$

$$G c_{p_b} (T_1 - T_b) + U_1 A_1 (T_w - T_b) - G_a c_{p_a} (T_b - T_a) - U_2 A_2 (T_b - T_a)$$

where:

$G$ = bed particle interchange rate, lb/hr
$Cp_b$ = bed particle specific heat, Btu/lb°F
$Cp_c$ = chamber wall specific heat, Btu/lb°F
$G_a$ = air flow through CBC, lb/hr
$Cp_a$ = air specific heat, Btu/lb°F
$U_1$ = heat transfer coefficient between fluidized CBC and interconnecting wall, Btu/ft$^2$hr°F
$A_1$ = area of interconnecting wall ft$^2$
$U_2$ = overall heat transfer coefficient between fluidized CBC and ambient air, Btu/ft$^2$/hr°F
$T_1$ = main bed temperature, °F
$T_o$ = inlet air temperature (initial temperature), °F
$T_w$ = interconnecting wall temperature, °F
$T_a$ = ambient air temperature, °F
$M_c$ = weight of CBC chamber, lb
$M_b$ = bed weight in CBC, lb
$t$ = time from fluidization, hrs
$T_c$ = chamber wall temperature °F

In turn,

$$M_c c_{p_c} \frac{dT_c}{dt} = U_3 A_3 (T_b - T_c) \tag{J-2}$$

where:

$U_3$ = heat transfer coefficient between fluidized bed and wall near temperature of fluidized bed, Btu/ft$^2$hr$^o$F
$A_3$ = area of chamber walls, ft$^2$

Combining Eqs. 1 and 2

$$M_b c_{p_b} \frac{dT_B}{dt} = \qquad \text{(J-3)}$$

$$G c_{p_b}(T_1 - T_b) + U_1A_1(T_w - T_b) - G_a c_{p_a}(T_b - T_b) - U_2A_2(T_b - T_a) - U_3A_3(T_b - T_c)$$

Since $T_c$, the temperature of the chamber wall, is not a constant and is related to $T_b$ in a complex manner, Equation 3 cannot be integrated. Despite this difficulty, it was found that the interchange rate, G, could be determined by evaluating Equation 1 at $T_b = T_o$. The result thus obtained also satisfied the equation

$$T_B = \frac{k_2}{k_1} - \left[\frac{k_2}{k_1} - T_0\right] e^{-\alpha t} \qquad \text{(J-4)}$$

where

$$k_1 = G c_{p_b} + U_1A_1 + G_a c_{p_a} + U_2A_2$$

$$k_2 = G c_{p_b} T_1 + U_1A_1T_w + G_a c_{p_a} T_c + U_2A_2T_a$$

$$\alpha = \frac{k_1}{M_c c_{p_c} + M_b c_{p_b}}$$

For the steady state case, t → ∞ the bed temperature should be given by

$$T_B = \frac{k_2}{k_1} \qquad \text{(J-5)}$$

RESULTS:

Table J-1 below gives the flow rate, G, obtained by evaluating Eq. J-1 at $T_b = T_o$, and $\frac{dT}{dt} = 0$ and Eq. J-5.

TABLE J-1. RESULTS OF INTERCHANGE EXPERIMENTS

| Test | Open Area, Sq. Inches | Indicated Flow Rate Pounds Per Hour Eq. J-1 | Indicated Flow Rate Pounds Per Hour Eq. J-5 | Flow Rate Per Unit Area, Eq. J-1 |
|---|---|---|---|---|
| A. | 8 | 690 | 570 | 12,400 |
| B. | 12 | 1,000 | 1,390 | 12,100 |

It is believed that the results of Eq. J-1 would be most reliable since fewer constants must be evaluated.

For larger openings, which would be used in commercial designs, a value of 10,000 lb/hr per square foot of opening is recommended.

This data may be roughly compared with the diffusion coefficient suggested by Argonne (ANL/ES/CEN-F017, page 29) for -1/16 inch particles diffusing in a bed fluidized at 3 ft/sec. The value given there was D = 300 $ft^2$/hr based on experiments at the Coal Research Establishment.

Assume

$$D = \frac{kd_p(V - V_{mf})}{V_{mf}} \qquad \text{(J-6)}$$

which has been given for a "packed-fluidized bed" where

$k$ = an empirical constant, ft/sec
$d_p$ = particle diameter, ft
$V$ = superficial gas velocity, ft/sec
$V_{mf}$ = minimum fluidization velocity, ft/sec

Comparing a -1/8 inch bed fluidized at 12 fps (HV) with a -1/16 inch bed fluidized at 3 fps (LV), a high-velocity diffusion coefficient may be estimated from

$$D^{HV} = D^{LV}\left(\frac{\bar{d}_p^{HV}}{\bar{d}_p^{LV}}\right)\frac{(V^{HV} - V_{mf}^{HV})/V_{mf}^{HV}}{(V^{LV} - V_{mf}^{LV})/V_{mf}^{LV}} \qquad \text{(J-7)}$$

where

$$D^{LV} = 300\ ft^2/hr;\ (\bar{d}_p^{HV}/\bar{d}_p^{LV}) = 2;\ V^{HV} = 12\ fps,\ V^{LV} = 3\ fps$$

$$V_{mf}^{HV} = 3\ fps,\ V_{mf}^{LV} = 1\ fps$$

Substituting these values into Eq. J-7, D may be evaluated

$$D^{HV} = 300(2)\left(\frac{(12-3)/3}{(3-1)/1}\right) = 900\ ft^2/hr$$

A "pseudo" diffusion coefficient may also be computed based on the data of the interchange experiments. The interchange rate, R, may be equated with a diffusion coefficient and concentration gradient

$$R\left(\frac{lb}{ft^2\,hr}\right) = D\left(\frac{ft^2}{hr}\right)\frac{dC\left(\frac{lb}{ft^3}\right)}{dX\ (ft)} \qquad \text{(J-8)}$$

Assume that dC describes the concentration of hot particles, i.e., the density of the expanded bed, = 50 lb/ft³.
Assume dX is the distance over which the hot particles travel, i.e., the distance between the point at which the main bed temperature is not depressed by the cold CBC interchange and the end of the CBC, dX ≃ 3 ft

$$R = 12{,}000\left(\frac{lb}{ft^2\,hr}\right) = D\left(\frac{ft^2}{hr}\right)\frac{50\ (lb/ft^3)}{3\ (ft)}$$

$$D = 720\ ft^2/hr \quad \text{(through a barrier)}$$

# APPENDIX K

## FBM-CBC TEST SUMMARY DATA

TABLE K-1. FBM-CBC TEST SUMMARY DATA

| Test Number | B-5 | B-7a | B-7b | B-8 | B-9* | B-10* |
|---|---|---|---|---|---|---|
| Coal Type | E.Ky. | Pike Cty. | | Perf. 8 | Pittsburgh #8 Seam | |
| % S | .9 | .9 | .9 | 3.3 | | |
| % Ash | 10.7 | 13.9 | 13.9 | 13.5 | | |
| Overall Heat Balance | | | | | | |
| Fuel heat input, kBtu/hr | 7470 | 7600 | 6800 | 7300 | 6600 | 7100 |
| Boiler steam gain | 3960 | 4060 | 3660 | 3760 | 3550 | 3760 |
| Circulated $H_2O$ absorption | 2226 | 2008 | 1887 | 1907 | 1764 | 1674 |
| Flue gas loss | 979 | 1253 | 964 | 1124 | 937 | 1134 |
| Combustible loss** | 265 | 344 | 279 | 402 | 422 | 406 |
| Radiation & Unaccounted | 100 | 52 | 20 | 123 | 57 | 124 |
| Thermal efficiency, % | 83 | 80 | 82 | 78 | 82 | 76 |
| Overall Combustion efficiency, % | | | | 94 | 94 | 94 |
| $CaCO_3$ feed rate, lb/hr | - | - | - | - | - | - |
| FBM | | | | | | |
| Air rate, lb/hr | 5550 | 7150 | 5890 | 5725 | 5470 | 5685 |
| Coal rate | 615 | 645 | 575 | 680 | 580 | 610 |
| Bed temperature, °F | 1580 | 1650 | 1620 | 1650 | 1600 | 1600 |
| Fly ash output, lb/hr | 160 | 194 | 174 | 201 | 172 | 170 |
| Fly ash carbon content, % | 59 | 54 | 54 | 54 | 53 | 51 |
| Flue gas concentrations: | | | | | | |
| $O_2$, % | 2.6 | 3.0 | 2.5 | 2.7 | 3.0 | 3.0 |
| $SO_2$, ppm | 650 | 660 | 680 | 3100 | 3300 | 3350 |
| NO, ppm | 480 | 420 | 430 | 400 | 450 | 540 |
| HC, ppm | 90 | 45 | 80 | 365 | 150 | |
| $C_b/C_{in}$ | | | | 76 | 76 | 78.4 |
| CBC | | | | | | |
| Air rate, lb/hr | 905 | 985 | 880 | 925 | 910 | 800 |
| Coal rate, lb/hr | - | 10.5 | - | - | - | - |
| Fly ash input, lb/hr | 150 | 184 | 166 | 190 | 165 | 160 |
| Bed temperature, °F | 1960 | 1950 | 1980 | 1960 | 2000 | 2050 |
| Fly ash output, lb/hr | 78 | 100 | 96 | 113 | 100 | 102 |
| Fly ash carbon content, % | 21 | 23 | 20 | 23 | 22 | 23 |
| Cell combustion efficiency, % | 82 | 75 | 78 | 75 | 74 | 72 |

TABLE K-1. (Continued)

| Test Number | B-5 | B-7a | B-7b | B-8 | B-9* | B-10* |
|---|---|---|---|---|---|---|
| Flue gas concentrations: | | | | | | |
| $O_2$, % | 5 | 6 | 5.4 | 6 | 5 | 4.5 |
| $SO_2$, ppm | 800 | 620 | 600 | 1100 | 1300 | 750 |
| NO, ppm | 570 | 490 | 510 | 520 | 480 | 440 |
| HC, ppm | 30 | 20 | 30 | 30 | 30 | 20 |

* Initial condition before additive addition.

** Includes C, ash heat, blowdown and HC.
$C_b/Cin$ = (Carbon burned / Carbon in Coal) x 100

TABLE K-1. (Continued)

| Test Number | B-11* | B-11b | B-12* | B-12b | B-12e |
|---|---|---|---|---|---|
| Coal Type | Perfect 8 | | Pittsburgh #8 Seam | | |
| % S | 3.3 | | | | |
| % Ash | 13.5 | | | | |
| Overall Heat Balance | | | | | |
| Fuel heat input, kBtu/hr | 7120 | | 7330 | 7500 | 7600 |
| Boiler steam gain | 3670 | | 3760 | | 3860 |
| Circulated $H_2O$ absorption | 1486 | | 1775 | | |
| Flue gas loss | 1002 | | 1190 | | |
| Combustible loss** | 626 | | 422 | | |
| Radiation & Unaccounted | 285 | | 129 | | |
| Thermal efficiency, % | 72.5 | | 76 | | |
| Overall Combustion efficiency, % | 89 | 90 | 93 | 87 | 84 |
| $CaCO_3$ feed rate, lb/hr | - | 118 | - | 124 | 215 |
| FBM | | | | | |
| Air rate, lb/hr | 5710 | | 5660 | | |
| Coal rate | 620 | 620 | 635 | 650 | 660 |
| Bed temperature, °F | 1620 | 1610 | 1640 | 1570 | 1600 |
| Fly ash output, lb/hr | 200 | 364 | 174 | 338 | 361 |
| Fly ash carbon content, % | 51 | 45 | 52 | 42 | 33 |
| Flue gas concentrations: | | | | | |
| $O_2$, % | 3.2 | | 2.8 | 3.8 | 2.4 |
| $SO_2$, ppm | 3450 | 1900 | 3200 | 1950 | 1700 |
| NO, ppm | 370 | 380 | 390 | 430 | 390 |
| HC, ppm | 120 | 75 | 125 | - | 70 |
| CB/CIN | 75 | 61 | 78.5 | 67 | 73 |
| CBC | | | | | |
| Air rate, lb/hr | 780 | | 820 | | |
| Coal rate | - | - | - | - | - |
| Fly ash input | 182 | 331 | 158 | 308 | 329 |
| Bed temperature, °F | 1980 | 1970 | 1950 | 1910 | 1900 |
| Fly ash output, lb/hr | 133 | 222 | 104 | 235 | |
| Fly ash carbon content, % | 33 | 18 | 26 | 24 | 24 |
| Cell cumbustion efficiency, % | 53 | 73 | 67 | 57 | 35 |

TABLE K-1. (Continued)

| Test Number | B-11* | B-11b | B-12* | B-12b | B-12e |
|---|---|---|---|---|---|
| Flue gas concentrations: | | | | | |
| $O_2$, % | 3.5 | 2.4 | 2.4 | 2.3 | 0.6 |
| $SO_2$, ppm | 1420 | 940 | 2500 | 2000 | 3950 |
| NO, ppm | 560 | 540 | 510 | 430 | 580 |
| HC, ppm | 20 | - | - | - | 35 |

* Initial condition before additive addition.

** Includes C, ash heat, blowdown and HC.

TABLE K-1. (Continued)

| Test Number | B-13c | B-14 | B-15 | B-16b | B-17b |
|---|---|---|---|---|---|
| Coal Type | | | | | Powhatan |
| % S | | | | | 4.5 |
| % Ash | | | | | 12.1 |
| Overall Heat Balance | | | | | |
| Fuel heat input, kBtu/hr | 7600 | 6750 | 6030 | 5472 | 11,065 |
| Boiler steam gain | 3810 | 3500 | 3050 | 2690 | 4365 |
| Circulated $H_2O$ absorption | 1699 | 1426 | 1388 | 1138 | 1807 |
| Flue gas loss | 1159 | 1535 | 1324 | 628 | 1469 |
| Combustible loss** | 439 | 107 | 104** | 780 | - |
| Radiation & Unaccounted | 466 | 182 | 167 | - | 1860 |
| Thermal efficiency, % | 73 | 73 | 73 | 70 | 56 |
| Overall Combustion efficency, % | [~93.3] | 99.1 | 98.6 | - | - |
| $CaCO_3$ feed rate, lb/hr | Bed | - | - | 240 | 140 |
| FBM | | | | | |
| Air rate, lb/hr | 5440 | 5830 | 4821 | 4699 | 8400 |
| Coal rate | 666 | 592 | 529 | 480 | 846 |
| Bed temperature, °F | 1600 | 1700 | 1590 | 1570 | 1640 |
| Fly ash out, lb/hr | 165 | 153 | 140 | 300 | 318 |
| Fly ash carbon content, % | 45 | 48 | 49 | 22 | 33 |
| Flue gas concentrations: | | | | | |
| $O_2$, % | 3.3 | 3.5 | 3.3 | 6 | 3.4 |
| $SO_2$, ppm | 1500 | 3100 | 2600 | 700 | 1700 |
| NO, ppm | 290 | 370 | 310 | 300 | - |
| HC, ppm | 200 | 70 | 20 | 0 | - |
| CB/CIN | 83.2 | 81.5 | 80.5 | 79 | 83 |
| CBC | | | | | |
| Air rate, lb/hr | 840 | 1350 | 1880 | 1194 | 0 |
| Coal rate | - | - | - | coal feed | - |
| Fly ash input | 150 | 139 | 127 | 0 | 0 |
| Bed temperature, °F | 1800 | 2000 | 2000 | 1860 | 170 |
| Fly ash out, lb/hr | 109 | 76 | 68 | - | 0 |
| Fly ash carbon content, % | | 4.3 | 7 | - | - |
| Cell combustion efficiency, % | [~60] | 95 | 93 | - | - |

TABLE K-1. (Continued)

| Test Number | B-13c | B-14 | B-15 | B-16b | B-17b |
|---|---|---|---|---|---|
| Flue gas concentrations: | | | | | |
| $O_2$, % | 1 | 8.5 | 10 | 9 | - |
| $SO_2$, ppm | 4500 | 1000 | 1800 | 1850 | - |
| NO, ppm | 350 | 460 | 470 | - | - |
| HC, ppm | 600 | 10 | 20 | 0 | - |

* Initial condition before additive addition.

** Includes C, ash heat, blowdown and HC.

# APPENDIX L. DISCUSSION OF FLUIDIZED-BED COMBUSTION VARIABLES

## A. COAL RELATED VARIABLES

The description of any particular coal may include a large number of chemical and physical parameters, some of general interest, others of interest only in a limited number of applications.

The following basic coal variables are presently believed to influence, to some extent, either the design or operation of a fluidized-bed combustor:

1. Moisture content
2. Ash content
3. Calorific value
4. Sulfur content
5. Ash fusibility
6. Volatile content

The first three variables determine the theoretical mass flow rate of fuel required to deliver the desired steam flow. Hydrogen content might also be added to this list.

The sulfur content determines the additive requirements.

Variables 5 and 6 "suggest" the upper and lower limits for operating temperature of a fluidized-bed combustor.

In addition to these properties, for which standard tests have been suggested, size consist, density and hardness of the ash is of significance to the fluidized-bed boiler designer and operator.

### 1. Definitions and Methods for Determination of Coal Related Variables

a. Moisture in coal is defined in ASTM D121-62 as, "Essentially water, quantitatively determined by definite prescribed methods which may vary according to the nature of the material". It is further noted the methods suggested may not determine all of the water present and that ASTM D271 describes the prescribed method for coal and coke. It may be assumed that the same methods will apply to fly ash or char.

   Within the limits of accuracy required for design and operation ASTM D271 adequately measures moisture in coal.

In this method a specially prepared sample is heated in dry air at 104 - 110°C for one hour and the moisture content is defined as the loss of sample weight. The method does not determine some of the water chemically bound which then reports as hydrogen in the ultimate analysis. An additional source of error is that due to oxidation weight gain by the coal in the heated air. An inert gas drying method has been tentatively suggested to overcome this difficulty.

Moisture may affect the combustion process in two ways. First, it may be assumed that ignition does not occur until essentially all moisture has been driven from the coal. However, heat-up time in a fluidized bed is small (ten's of milliseconds) when compared with time spent in the bed (hundreds to thousands of milliseconds).

The second manner in which moisture affects the combustion process is in the energy removed from the combustion bed in heating, vaporizing and superheating water. Assuming a partial pressure of 1 psi for water vapor in the flue gas a pound of water in the fuel removes ∿ 1800 Btu from a 1600°F bed. This effect may be anticipated in the design, if the moisture content of the coal to be burned is known, or in changes in appropriate operational parameters (bed height primarily) if the source of coal is changed.

b. Ash in coal is defined in ASTM D 121-62 as the "Inorganic residue remaining after ignition of combustible substances, determined by definite prescribed methods". In addition it is noted that ash so determined may not be identical, in composition or quantity, with the inorganic substances present in the material before ignition. It is further noted that the prescribed method for ash content of coal is that of ASTM D 271.

Within the limits of accuracy required for the design and operation ASTM D 271 adequately measures the quantity of ash in coal. Except for calcium oxide content the chemical composition, per se, of ash is of no interest to the designer or operator of a fluidized-bed boiler. Where increased accuracy is desired as to the quantity and composition corrections may be made using the methods of Parr and Rees.

The effect of ash content on the combustor is similar to that of moisture in that the ash removes energy from the bed. One pound of ash elutriating from the bed at 1600°F removes approximately 400 Btu.

Ash in coal, as fired in a fluidized-bed boiler, may be retained in the bed or may be carried out with the gas. The retention or expulsion depends on the nature of the ash, primarily its particle size, density and hardness.

Coal ash is believed to have originated in two ways. First, some ash is that mineral matter absorbed by the vegetation which gave rise to coal. Second, the bulk of the ash arose from the silt washed in as the precursors of the coal beds were forming. It is the sedimentatous matter which would be expected to provide the particles which might be retained by the bed. A final source of material which is defined as ash are impurities taken with the coal in mining, rocks or earth, and impurities which intrude the coal in handling and shipping i.e., railroad bed gravel, rust and scale, nuts and bolts, etc.

No standard tests have been devised for the property of coal ash which affects a fluidized-bed boiler since it is superfluous to conventional firing methods.

c. The calorific value or heating value of the fuel is determined following the methods of ASTM D 271. In this method a sample of the coal is burned with oxygen in a bomb calorimeter following the prescribed procedure.

The value obtained in this procedure is the gross or higher heating value of the fuel burned at constant volume with the moisture in the fuel and water vapor resulting from the hydrogen in the fuel being cooled to about 20°C.

In an actual fluidized-bed boiler the fuel will be burned at constant pressure and the gases discharged at above 100°C. Two corrections are therefore required, one to account for the combustion mode and the second to account for the discharge of water vapor. The latter correction is made in the procedures prescribed for determination of boiler efficiency wherein the accepted convention is to charge the boiler with the enthalpy contained in the discharged water vapor. The correction for constant pressure combustion. generally small, is derived at the close of this section.

The effect of calorific value of the coal may be treated in a straightforward manner in the design and operation of a fluidized-bed combustor.

d. Sulfur in coal is defined as the total elemental sulfur equivalent of the organic and inorganic sulfur forms. ASTM D 271 prescribes the methods for determining this sulfur. ASTM D 2492 describes the methods for differentiating between forms of sulfur; sulfate, pyritic and organic.

Sulfate sulfur in coal is of some interest in that it may be retained in the ash after combustion in a fluidized bed. However, the amount of sulfate sulfur is generally small and its retention or release as $SO_2$ does not effect the design and operation of a fluidized-bed boiler.

The sulfur content of coal has been treated as a control variable and was found not to affect significantly the stoichiometric limestone requirement within the range evaluated, i.e., 2.6% S to 4.5% S.

e. Ash fusibility denotes the temperatures at which coal ash deforms and flows when heated in certain atmospheres. The tests which are empirical are described in ASTM D 1857.

Four temperatures are noted: An initial ash deformation temperature (IT) at which the apex of the specially prepared ash cone rounds; a softening temperature (ST) at which the cone is spherical and height equals base width; a hemispherical temperature (HT) at which the cone is hemispherical and the height is one half the base width and a fluid temperature (FT) at which the cone height is 1/16" or less. These four temperatures are noted when the test furnace atmosphere is reducing and also when it is oxidizing. Eight values are therefore normally provided for any coal.

It has been empirically determined that it is possible to operate a stable coal-fired fluidized bed at the softening temperature (ST) of the coal ash as determined in a reducing atmosphere.

This appears to establish the upper limit for operation of such a system.

Although the empirical limit appears to exist at ST (reducing) a correlation of the bed collapse temperature and the fusibility of coal ash has not been developed.

f. Volatile matter is defined in ASTM D 121 as "those products, exclusive of moisture, given off by a material as gas or vapor, determined by definite prescribed methods which may vary according to the nature of the material". It is noted that for coal and coke the prescribed method is that of ASTM D 271. Although the amount of weight lost in heating coal in a fluidized bed will differ from that lost following ASTM D 271 it has been determined empirically that a correlation exists between ignition temperature and volatile content as determined in the ASTM test.

Ignition temperature is defined as the temperature at which more heat is generated than is lost to the surroundings. It is a function of both fuel properties and reactor design.

The correlation between ignition temperature and volatile matter although not quantified is sufficiently well understood for purposes of fluidized-bed boiler design and operation.

A similar correlation exists for ignition stability temperature, a temperature somewhat higher than the ignition temperature. This too, although not quantified is sufficiently well understood for design purposes.

Although the temperature at which coal ignites and burns stabily in a fluidized bed is related to its volatile content other tests are required if predications of the reaction rate are desired. These tests would, by necessity, be empirical and conducted in a test system almost identical to the fluidized-bed boiler.

One test procedure has been suggested in which a "burning profile" is plotted for various coals and is cited as being capable of predicting carbon loss from a pulverized-fuel boiler. This test however omits the comminution which occurs in the fluidized bed.

## B. COMBUSTION AIR RELATED VARIABLES

Air for combustion is, by convention, considered to consist of "dry air" and water vapor. Dry air consists, again by convention, of oxygen (21.0% by volume) and a mixture of inert gases which are considered nitrogen (79.0% by volume). The pressure and temperature of the air also influence design and performance.

### 1. Water Vapor

The water vapor content may be expressed in a number of ways but must be reduced to pounds of water vapor per pound of dry air (specific or absolute humidity) for use in a combustion calculation. Determining the wet and dry bulb temperature with a sling psychrometer and the barometric pressure provides data of sufficient accuracy for calculating the effects of water vapor content.

The water vapor content of the combustion air affects bed temperature and boiler efficiency in predicable ways. A pound of water vapor entering the bed at 600°F and leaving at 1600°F removes approximately 522 Btu from the bed. A pound of water vapor entering the air heater at 80°F and exiting with the flue gas at 350°F is responsible for an energy loss of approximately 120 Btu. A precise value for efficiency loss due to moisture in combustion air may be derived from a calculation of enthalpies at the precise partial pressures and temperatures. The convention, however, is to assume the exiting water vapor at 1 psia.

### 2. Pressure and Temperature

Combustion air becomes available to the boiler system at ambient or atmospheric conditions. In order to make it available to the combustion process, it must be compressed and accelerated. In passing through the air distributor and the bed, some of the potential energy in the air is transferred to the bed. The effect on bed temperature is small, although stagnation temperature equations could be used to calculate the effect of various grid designs and bed depths on the bed energy balance.

Combustion air temperature is an important parameter since the inert components and unreacted oxygen remove energy from the bed. One pound of nitrogen entering the bed at 600°F and exiting at 1600°F removes approximately 271 Btu from the bed. Unreacted oxygen has a similar effect, 254 Btu/lb required to go from 600°F to 1600°F. Use of air in excess of that required for complete combustion acts to cool the bed in a way generally predicatable on the basis of the energy rise of these two gases and the water vapor content.

### 3. Discussion

The combustion air related variables are design variables, not control variables. The design convention is to assume the unit 0 feet above sea level, air at 80°F and 60% relative humidity. In designing a unit for a specific installation, the user must specify local conditions. Although a slight increase in oxygen partial pressure is obtained under a potential mode of operation, i.e., pressurized vs. balanced draft, its effect on combustion intensity could not be measured.

Combustion air temperature is also a design variable. Air passing into the bed achieves bed temperature so near the bottom of the bed that the effect of inlet air temperature alone on combustion is probably not detectable.

## C. BED MATERIAL RELATED VARIABLES

### 1. General

The bed material of a fluidized-bed boiler is defined as that inert, solid, granular substance which makes up the bulk of the fluidized bed. The gases, burning carbon particles, and dusts (limestone or fly ash) are excluded from this definition. The bed material supplies ignition energy to the incoming coal, transfers energy to the boiler tubes, dilutes the burning coal so as to prevent clinkering, provides an effective radiator to the freeboard and convection bank, attrites burning coal particles and may act to "accept" sulfur dioxide. It also may decrepitate and require replacement, erode the boiler tubes and air distributor and change with time in a way that disturbs the process.

### 2. .Source or Name

It is necessary to first identify the material as to its source. Materials being studied today include: natural coal ash, sintered coal ash, alumina, limestone, zirconia and refractory grog.

### 3. Particle Density

Particle density effects fluidization velocity requirements, elutriation rate and heat transfer properties.

For most substances that would be suitable as a bed material, the pycnometer methods of ASTM Standards D153-54, D792-60 and D1480-62 will be appropriate.

Since particle density is temperature dependent, it is necessary to correct the density found at room temperature to the temperature at operating conditions.

The pycnometer method requires wetting the material with a liquid, normally water. For materials which will react with water, such as calcined limestone, an organic liquid, such as carbon tetrachloride or hexane, should be used.

The particle density which is significant to fluidization and elutriation phenomena will probably be that determined with a fluid which did not wet the particles or drive air from any pores. Viscous or a non-wetting liquid, such as mercury, might be used here.

A factor such as shape, which influences the particle behavior profoundly, is so poorly understood that small errors in particle density description cannot be considered serious. The use of a wetting liquid, appropriate to the chemical nature of the particle, provides an adequate measure of particle density.

## 4. Bulk Density

The fan power requirements are a strong function of the total bed weight. Bed weight is in turn proportional to the bed depth and its bulk density. Settled bed density is also an appropriate term for this parameter and is probably more meaningful. The methods prescribed in ASTM D291-60 will provide data of sufficient accuracy for design purposes.

## 5. Size Consist or Particle Size Distribution

The behavior of the operating bed appears to be strongly influenced by the particle size distribution or size consist. Certainly, the ability to turn down without any portion of the bed becoming static is related to the size consist. Again, particle shape which is poorly understood appears to be of equal importance.

The methods of ASTM C136-67, a test for aggregates, appear adequate for particle size distribution. A mean or average particle diameter may then be derived using a definition by Reboux.

$$\frac{1}{D_p} = \sum \frac{X}{d_p} \qquad \text{(L-1)}$$

where

$D_p$ = average or mean particle diameter

X = weight fraction in a certain narrow cut size range

$d_p$ = simple mean diameter in the narrow cut

Generally, the particle size distribution and average particle diameter have been ignored in favor of a specification of the top particle diameter which is simple to establish. Top size is useful since it clearly differentiates between the high velocity and low velocity systems. A specification of both top size and mean size is probably required for a good description.

### 6. Particle Shape or Rugosity

Natural ash beds tend to be made up of particles which are irregular flat discs. Most other materials tend toward spheres. Rugosity is the ratio of the actual external surface area to the surface area of a hypothetical sphere. A shape factor has been defined for use in various correlations but appears not to be very useful and is very tediously determined. If fluidized-bed boilers come into widespread use, an investigation of particle shape on performance may be justified.

### 7. Friability or Resistance to Abrasion

Bed particles impact heat transfer surfaces, the air distributor and one another and are constantly being diminished in size. Although the only conclusive test is at temperature and in a device similar to the boiler, methods are available for gross screening of materials as candidate substances. The methods of ASTM D441-45, Tumbler Test for Coal, or ASTM C131-66, for aggregates, describe adequate procedures and apparatus which might be adapted for use as a standard method for fluidized-bed materials. Where the material to be evaluated is natural coal ash, a sample of coal would be ashed at low temperatures to preclude sintering and tested. Limestones would be difficult to test properly since, in operation, they would be calcined and partially sulfated.

### 8. Loss-on-Charging

Bed materials other than natural ash would normally be purchased on a weight basis (or by a volume measure, cubic yards, which is essentially equivalent). It may be anticipated that the as-received particle size distribution will differ from that required in the boiler unless a double screened, and therefore more costly, material is specified.

The selected material may suffer a loss on ignition due to organic component, moisture, and carbonate or water of hydration. In preparing a material for charging by crushing, fines are produced which will elutriate on charging to the boiler. On charging, the sharp edges attride rapidly and additional mass is lost.

Although no standard test exists for "loss-on-charging," it might be constructed from existing tests for loss-on-ignition, resistance to abrasion and screen analysis.

Certain potential bed materials may also decrepitate on charging due to the rapid heating. Some limestones or materials with water of hydration might literally explode on charging. It appears that actual use in a fluidized-bed boiler is the most meaningful test of "loss-on-charging."

### 9. Discussion

The choice of bed material appears to be a design variable. It is doubtful that bed material affects the combustion process to any significant degree. Although not listed above, it should be obvious that a high melting point, inertness and low cost are basic requisites of any candidate material.

## D. BED TEMPERATURE

The temperature of an oxidizing fluidized bed has been defined as the temperature indicated by a thermocouple immersed in the bed. While the precise type of couple and its heat loss by conduction may affect the temperature indicated, errors above 5% are unlikely.

The temperature so indicated represents the average temperature of the energy sources and sinks which make up the bed. It has been assumed that this average temperature closely approximates the average temperature of the inert particles which make up the bulk of the bed.

It is possible by use of thermocouples to detect the macroscopic gradients, both vertical and horizontal, which result from the combined effects of combustion and heat loss gradients.

It is not possible by means of thermocouples to determine the microscopic gradients which must exist due to second law considerations. The maximum temperature to which a particle has been exposed might be determined by analysis of samples removed from the bed.

While bed temperature is a response variable, it is actually one of the fundamental design parameters. Bed temperature has been found to affect combustion intensity, carbon carry-over, $SO_2$ removal by limestone, and NO generation.

## E. FUEL ENERGY INPUT RATE

The fuel energy input rate is defined as the higher heating value of the fuel (Btu/lb) multiplied by the fuel rate (pounds/hr or pounds/hr for each square foot of bed surface). When the fuel enters the combustor at a temperature above the reference temperature, usually the ambient air temperature or 80°F, the sensible heat in the fuel is added. By convention, coal is assumed to have a mean specific heat of 0.3 Btu/lb°F. More precise values for specific heat may be used although the fraction of the total energy carried into the combustor as sensible heat in the raw fuel is very small in any real case. When the fuel is recycled particulate, the sensible heat may be important.

Fuel energy input rate, coal rate, fuel rate and firing rate are synonymous terms. While this control variable is related to the heat release rate or combustion rate (a response variable), it is not identical. Not all the fuel entering the bed actually burns (10-20% of the energy in the fuel may appear as unburned combustible), and not all of the fuel which burns in the combustor actually burns in the bed.

## F. SUPERFICIAL VELOCITY

The specification of an approximate superficial velocity in a fluidized-bed reactor is useful in distinguishing between kinds of fluidized beds, that is, true fluid beds and what has been termed "teeter"* beds. Superficial velocity is normally considered a control variable. In fact, it is response variable. Superficial velocity is defined as the velocity of the fluid which would exist at a specific point in the reactor if it were empty of both particles and internals. The specific point which is usually designated is the top of the expanded bed. It may be specified knowing the total mass flow rate, the specific volume at temperature and the cross sectional area of the reactor at the point of interest. It may be useful as a tool in determining, for fluid beds, from correlations available in the literature, minimum fluidization velocity, bed expansion, elutriation rates and particle residence times. Unfortunately, many of the correlations proposed for the fluid bed do not hold for teeter beds.

For a combustion system, which may be packed with internals and in which all the combustion does not occur in the bed, specification of a superficial velocity is of relatively little value as a design tool.

In lieu of superficial velocity, it would be more appropriate to specify the air rate and the average bed temperature. If the superficial velocity is of interest, an approximate value may be computed from a knowledge of these two parameters.

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*Beds made up of large particles ( + 20 mesh) permitting high mass gas flows

## G. CARBON LOSS

Comparing the combustion conditions of a fluidized-bed boiler with a pulverized coal furnace, bed temperature and oxygen partial pressure in the fluidized bed are lower and coal particle sizes greater. As a result, it has not been found possible to burn all of the coal as it is fed to the bed. Some particles leave the bed and are carried out of the region in which burning may occur with the flue gas before combustion is complete.

The potential loss of fuel value due to carry-over of this unburned particulate combustible matter has been termed "carbon loss" since the major combustible component is carbon. The particles, however, still contain some hydrogen which increases its heating value. While the use of the term carbon loss is a useful shorthand, since it describes the major potential inefficiency of fluidized-bed boilers, it is more appropriate to specify the combustion efficiency as defined below:

$$CE = 1 - \frac{H_o}{H_i} \qquad \text{(L-2)}$$

where

$CE$ = combustion efficiency, dimensionless

$H_o$ = heating value of unburned fuel components leaving, etc., Btu

$H_i$ = heating value of fuel fed to system, Btu

The term $H_i$ may be considered as the integrated higher heating value of fuel fed over a specified period.

The term $H_o$ is the sum of all combustible matter leaving the system. It would include the heating value of:

a. Combustible in bottom ash, primarily carbon
b. Combustible in fly ash, primarily carbon
c. Combustible in flue gas as hydrogen, as carbon monoxide and as hydrocarbons

While a precise determination of the heating value of the solid matter and gas might be preferable, conventional practice has been to assume the following values:

solid combustible = 14,500 Btu/lb
carbon monoxide = 10,160 Btu/lb
hydrogen = 319 Btu/ft$^3$ (STP)

No convention exists for the heating value of hydrocarbons, but it might be assumed that the hydrocarbons as reported by a flame ionization chamber, as methane, do describe the gas. Therefore, the heating value, the psuedo methane, might be taken as 1030 Btu/ft$^3$ (STP).

## CORRECTION OF CONSTANT VOLUME HEATING VALUE TO CONSTANT PRESSURE HEATING VALUE.

Nomenclature:

C = Specific heat, Btu/lb °R

H = Weight of hydrogen in fuel, %

J = Mechanical equivalent of heat = 778.26 ft lb per Btu

M = Pound moles of gaseous constituents per pound of fuel

N = Weight of nitrogen in fuel, %

P = Pressure, pounds per $ft^2$

Q = Quantity of heat absorbed by system from surroundings, Btu/lb of fuel

R = Universal gas constant = 1545 ft lb/lb mole °R

T = Calorimeter temperature, °R

U = Internal energy, Btu

V = Volume, $ft^3$/lb of fuel

Δ = Net change in quantity following

Subscripts

v = constant volume

p = constant pressure

1 = start of process

2 = end of process

(l) = liquid

(s) = solid

(g) = gaseous

The derivation of the correction is based on the first law of thermodynamics and the perfect gas law. In burning a sample of coal in a bomb, no work is performed and there is no kinetic energy change and the first law may be expressed as

$$Q_v = U_2 - U_1 = C_p \Delta T - \Delta PV/J \qquad \text{(L-3)}$$

But

$$C_p \Delta T = Q_p \qquad \text{(L-4)}$$

Rearranging

$$Q_p = Q_v + \Delta PV/J \qquad \text{(L-5)}$$

The perfect gas law states,

$$\Delta PV = \Delta MRT \qquad \text{(L-6)}$$

Therefore,

$$Q_p = Q_v + \Delta MRT/J \qquad (L-7)$$

To determine the ΔM, the change in the number of moles of gaseous constituents, the combustion reactions in the calorimeter are examined.

$C\ (s) + O_2\ (g) \rightarrow CO_2\ (g)$,

1 solid mole + 1 gaseous mole → 1 gaseous mole, no net change

$H\ (s) + \frac{1}{2}O_2\ (g) \rightarrow H_2O\ (l)$

1 solid mole + ½ gaseous mole → 1 liquid mole, -½ gaseous mole

$S\ (s) + O_2\ (g) \rightarrow SO_2\ (g)$

1 solid mole + 1 gaseous mole → 1 gaseous mole, no net change

$N_2\ (s) \rightarrow N_2\ (g)$

1 solid mole → 1 gaseous mole, + 1 gaseous mole

Other fuel constituents are assumed to undergo no net change, i.e.,

$H_2O\ (l) \rightarrow H_2O\ (l)$

$O_2\ (s) \rightarrow O_2\ (s)$

$Ash\ (s) \rightarrow Ash\ (s)$

$$\Delta M = \frac{-N}{28.016} + \frac{\frac{1}{2}H}{2.016} \qquad (L-8)$$

Inserting into Eq. (L-7)

$$Q_p = Q_v + \left(\frac{H}{4.032} - \frac{N}{28.016}\right)\frac{RT}{778.26} \qquad (L-9)$$

Examples:

$N = 1\%$

$H = 4\%$

$T = 68°F$

$Q_v = 13{,}000$ Btu/lb ;

$$\frac{H}{4.032} = \frac{.04}{4.032} = .00993$$

$$\frac{-N}{28.016} = -\frac{.01}{28.016} = -.00036$$

$$\Delta = .00957$$

$$Q_p = 13{,}000 + .00957(1050) = 13{,}010 \text{ Btu/lb}$$

# APPENDIX M. WHY THE CARBON BURNUP CELL?

## M.1 Perspective

While the body of the report discusses work completed in October 1970, this section must properly be written to reflect what is known in January, 1972.

To answer, "Why the Carbon-Burnup Cell" it is necessary to explain, a bit more fully, what a fluidized-bed is, and how particles leave the bed, how coal burns and how coal burns in a fluidized-bed, why a fluidized-bed boiler is put together the way it is and why all of these factors could have led to the invention of the Carbon-Burnup Cell.

## M.2 Why Particles of Coal Leave a Fluidized Bed

A fluidized-bed was defined in Section 4.1.2 and in Section 4.1.3 it was noted that some particles added to, or made, in the bed, could be too large and would sink to the bottom of the bed. Other particles added to, or made, in the bed are too small to remain in the bed and are carried out and away from the bed by the gases which fluidized the bed or passed through the bed.

In practical fluidized-beds, some gas passes through the bed, either as bubbles or in channels, without participating in the fluidization of the particles. These channels or bubbles may cause the bed to be violently agitated. And, just as a violently boiling pot of water throws off large droplets of water, the violently fluidized-bed shoots up, into the space above the bed, even the largest particles which make up the fluidized-bed. The support of the upward flow of gas and the kinetic energy of the particles, imparted as they left the bed, tend to keep the particles moving upward.

Gravity acts to return the particles to the bed.

The height of the freeboard; i.e., the space above the bed, determines which of the large particles will finally lose their initial upward momentum and fall back into the bed.

If the chamber, which holds the fluidized-bed in its bottom, is tall enough, every particle which is too heavy to be supported by the upward flow of gas will fall back into the bed. If the pipe or duct which allows the gas to leave the chamber is too close to the top of the fluidized-bed; i.e., there is little freeboard, many of the large particles will be swept into the outlet pipe and will not fall back into the bed.

The small particles, whose terminal velocity is less than the actual gas velocity, cannot be made to return to the bed unless they are first separated from the gas in a dust collector. Fine particles appear in the bed because: (a) they are fed to the bed as fine particles, (b) small pieces are broken from large pieces, (c) large particles are consumed by chemical reactions until the particles become small, (d) chemical reactions change gases into fine particles. The large particles can be made to return to the bed by providing a baffle in the freeboard which the particles may strike, thus losing their initial upward momentum. These large particles cannot be supported, solely, by the upward flow of gas and so fall back into the bed.

Particles of coal, like the other particles which make up the fluidized-bed, will appear in the freeboard either because they are too small to remain in the bed or because they are thrown out of the bed by the violent agitation.

From what has been said above it should be clear that if particles of coal, fed to the bed, have not been totally consumed before they come near the top of the bed, some coal particles will appear in the freeboard. It should also be clear that the freeboard is an integral part of a fluidized-bed system. It also follows from what has been said that unless the coal particles return* to the bed, or are totally consumed in the freeboard, unburned fuel values will leave the furnace. These matters will be discussed in following sections.

Despite a greal deal of effort by researchers around the world, no general equations have been derived which can be used to predict how much material will enter the freeboard above a fluidized-bed. Some of the major chemical and petroleum companies have developed useful prediction tools for some types of fluidized-beds made up of fine particles which tell them how tall a vessel (i.e., its disengaging space) should be. These tools, carefully guarded trade secrets, are of no relevance here. No one, to our knowledge, has developed a prediction tool for the high gas velocities used in fluidized-beds made up of the large particles used in a fluidized-bed boiler.

## M.3 Coal Combustion

It was shown, in the preceding paragraphs, that the entire size spectrum of bed particles will appear in the freeboard, and under some conditions, in the outlet pipe, of a fluidized-bed boiler's furnace. Some of these will be coal derived particles which are commonly referred to as "char".

The material which follows is taken, in part from

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*e.g., by internal cyclones with dip legs.

Reference 12, sent by the OAP Project Officer in December 1971. Its useful explanation of pulverized coal (p.c.) combustion, with changes necessary to emphasize the differences between fluidized-bed combustion and p.c. combustion, has been used.

The coal particle undergoes complex physical and chemical changes during combustion which may be characterized roughly by the sequence of preheat, ignition and char burnout. The details, and time taken, in each phase will vary greatly among different flames as a consequence of differences in coal type, the coal particle size spectrum, the aerodynamic patterns in the combustion zone and beyond, and the temperature at which the burning occurs.

A number of the variables applicable to pulverized coal combustion have been studied in a pilot-scale furnace at IJmuiden by the International Flame Research Foundation. While fluidized-bed combustion is not identical to pulverized coal combustion, they both involve burning in a suspension, and the same basic processss.

After injection into a furnace, the coal particles are heated both by radiation and by convection. Time for temperature equibriation in a conventional P.F. furnace for 1 µm particles is about $10^{-6}$ seconds, $10^{-2}$ seconds for 100 µm particles and 10 seconds for 1000 µm particles.

For bituminous coal with a high volatile content, ignition is achieved rapidly and the volatiles are released and burn as a diffusion flame in fuel rich pockets. It is possible that sufficient fuel gas is released in the vicinity of a coal feeder of a fluidized-bed boiler to form bubbles which travel up through the bed with oxygen diffusing into the bubble.

For large particles, volatiles may be released at a rate high enough to prevent oxygen from approaching the developing char until several seconds after the particle enters the furnace.

The burnout of the char is relatively well understood, occurring by the diffusion of oxygen to the surface, reaction at the surface to carbon monoxide, and subsequent gas phase oxidation of the outwardly diffusing carbon monoxide to carbon dioxide.

The rate of char burning depends on the rate at which oxygen can approach the surface, the reaction rate of oxygen with carbon and the rate of escape of the carbon monoxide. The very thin boundary layers associated with fluidized beds aid char burnup.

The relatively low temperatures and large particles typical of a fluidized-bed boiler act to reduce the reaction rate at the surface and also to slow the mass transfer. Times to burnout for large particles of char in fluidized beds may be on the order of tens or even hundreds of seconds.

## M.4 Combustion of Coal in a Fluidized-Bed

The chemical history of a coal particle in a fluidized-bed boiler is similar to, although not identical to, a particle in a pulverized coal flame. Both methods of burning keep the fuel in suspension. In the fluidized-bed, the suspension is stationary as long as the particles are too large to be swept out of the bed. Particles of char which leave the bed experience the same phenomena as the char particles in a pulverized coal furnace, except that the relatively low temperature in the freeboard of the fluidized-bed boiler is less conducive to high burning rates.

If the burning process is either reaction-rate or desorption-rate controlled, as it appears to be, an increase in absolute pressure will not decrease the burnout time. Experiments reported by Hoy (13) confirm this.

From what has been described to this point, it should be clear that combustion of the solid portion of coal residue, the char, is relatively slow; too slow to be completed within the fluidized-bed before some of the particles are swept or thrown out of the bed into the freeboard. It would be possible to provide conditions in the freeboard which would permit the combustion of char to be completed. That these conditions are incompatible with the design goals of a fluidized-bed boiler is shown below.

## M.4 Why a Fluidized-Bed Boiler Is Put Together the Way It Is

In a pulverized-coal boiler, the firing rate rarely exceeds about 20,000 Btu/hr per cubic foot of furnace volume. In a fluidized-bed boiler, the rate can easily be ten times as great. The main cause of this difference is that in the fluidized-bed a heat transfer medium is provided and it is possible to remove the heat generated by the burning coal with tubes that are, in a sense, immersed in the flame. The ratio square feet of heat transfer surface per cubic foot of heat release volume, is 10 to 15 times higher in the fluidized-bed boiler than in the pulverized coal furnace.

One of the first test systems built by Pope, Evans and Robbins provided the desired surface-to-volume ratio by placing boiler tubes uniformly in the combustion space. The problems found in that design have been described in Reference 4. A design concept then evolved in which heat-transfer surface was banked so that the tubes then formed natural barriers and repeating elements. The resulting configurations was termed a "unit cell". A unit cell contained a fuel supply and sufficient heat transfer surface (boiler tubes) to absorb a major fraction of the energy released within the cell. Since tubes were not arranged close to and directly above the fuel outlets, the problem of erosion due to high velocity jets was avoided. It was also possible for a man to enter a cell and make repairs without dismantling the boiler. The cell lent itself to the development of effective light-off and turndown techniques, as well as facilitating shop fabrication of the boiler moduler. And most important here, cells could be used to perform specialized functions.

All Pope, Evans and Robbins design concepts for fluidized-bed boilers contain cells, and some designs proposed by others appear to contain cells. Two names have been given to this basic concept, which is such a radical departure from both conventional boiler practice and what might be called conventional fluidized-bed practice. The names to be used are: Pope-Bishop Boiler or Multicell Fluidized-Bed Boiler.

The modern pulverized coal furnace exceeds 90% combustion efficiency, not because there is an over-riding economic incentive to do so but because the furnace size is set by heat transfer needs of the process; the very high combustion efficiency is an "extra" that follows from this large furnace size. A fluidized-bed boiler furnace could also be designed to provide 99+% combustion efficiency. For example, the following could be done: (a) the size of the fluidized-bed boiler's furnace could be increased by a factor of ten, and (b) the operating temperature could be increased from 1,500°F, now considered adequate, to about 2,200°F, and (c) the excess air could be increased from about 15% to the 50% common in stoker fired units and (d) carbon-bearing fly ash could be recycled back to the furnace or in lieu of (a) to (d), (e) the air rate could be reduced so that the superficial velocity dropped from 15 feet per second, now used, to 0.5 feet per second at which carryover is negligible (14). Why these approaches to complete single-stage combustion are undesirable is discussed shortly.

A fluidized-bed boiler is put together the way it is because it is possible to build an extremely compact primary cell ("furnace") if it is permissible to solve the problem of fuel carryover by an adjunct means such as the Carbon-Burnup Cell. The sections which follow will outline some of the unsuccessful efforts at reducing loss of fuel as high-carbon fly ash by more direct techniques.

To close this section, it is appropriate to note here that the problem of fuel loss from a fluidized-bed boiler has been considered in "The Federal Research and Development Plan for Air-Pollution Control by Combustion-Process Modification" by Battelle Memorial Institute (15). The Battelle plan recommends, on page III-34, basic research which ". . .should permit identification of conditions under which CO can be kept to levels not exceeding 0.1 percent and carryover of unburned fuel can be lowered to the equivalent of carryover in existing combustion systems." The work to be described in this report indicates that the goals set in the Battelle plan have been met without requiring a basic alteration in the configuration of the fluidized-bed boiler.

## 4.6 Methods for Reducing Fuel Loss Considered Prior to Invention of Carbon-Burnup Cell

A number of techniques for reducing fuel loss from a fluidized-bed boiler were considered, and some tested, before the invention of the Carbon-Burnup Cell. None provided the degree of combustion efficiency desired without at the same time possessing an over-riding liability. These are discussed below under the heading Time, Temperature and Turbulence, the three factors known to increase the rate of combustion. The discussion is qualitative and brief.

### M.6.1 Time

(a) We learned in May 1965 (14) that studies at the Marchwood Laboratories of England's Central Electricity Generating Board (C.E.G.B.) indicated that at a superficial velocity much above 0.5 feet per second, carryover of char would be excessive in beds up to three feet deep.

Operation at low velocities was judged unfeasible, although Pope, Evans and Robbins conducted one low velocity test in a bed composed solely of fly ash.

(b) Increasing the depth of the bed, which would also act to increase the time for char particles to be burned, was considered and rejected. The power required by the high pressure blowers needed to fluidize a deep bed, over about three feet, would cause a fluidized-bed boiler to be uneconomical. There was also no evidence that such deep beds would provide the desired burnout.

CEGB's results in a three-foot bed at 0.5 fps might be duplicated in a 90 foot deep bed at 15 fps. This calculation is an oversimplification but the implication may be correct.

(c) The volume of the freeboard could not be increased to the extent required for complete combustion (at least a factor of 10) without losing one of the major advantages of the fluidized-bed boiler; i.e., compactness.

(d) Recycle of collected fines* back to the bed from whence they initially arose was tested a number of times at high superficial velocities and found unsatisfactory. Experiments at low superficial velocities, 2-4 fps, indicates that this technique will work (16). However, operation at 2-4 fps is uneconomical. In addition, if the dust collectors used for recycle are relatively efficient, an extremely high dust load is imposed on the system. If the collector is 90% efficient, the quantity of circulating dust may exceed, by a factor of ten, the ash being fed with the coal.

(e) In one test, collected fines were pelletized with coal fines into 1/2" lumps resembling charcoal briquettes. Although the loss of fuel was reduced by the increase in residence time, the improvement did not justify the complication of a pelletizing operation.

(f) Theoretical studies had indicated that a pulsing air supply could be used to pass relatively large quantities of air through a fluidized-bed without excessive carryover. One test was made of this approach but the results were not promising. Coal feeder plugging might result.

The conclusion drawn from this work and study was that no feasible method existed for increasing the time that the particles of char would spend in the furnace, at least to the extent required.

### M.6.2 Temperature

(a) Goldberger has shown (17) the effects of temperature on combustion efficiency of coal fed to a fluidized-bed combustor. It was shown that for a bed depth of about three feet, operated at 4-7 fps, feeding relatively fine coal, a temperature above 1900°F was required to achieve complete combustion. Goldberger did not study coarse coals or propose a method for feeding fine coals to a large diameter bed, since these did not require consideration for the application proposed by the sponsors of his work.

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*Which must compete with fresh coal for available oxygen.

(b) Tests by Pope, Evans and Robbins at bed temperature up to 2,100°F with coarse coals, shallow beds and high superficial velocities did not indicate that 99+% combustion efficiency could be achieved by operation in this mode. We also determined that high bed temperatures were inconsistent with sulfur-oxide control using limestone.

(c) The use of a two-stage combustor was considered. Coal would be fed to a fluidized-bed operating at, say 1,500°F, at superficial velocities exceeding 15 fps. The freeboard would be large and refractory lined. Sufficient fuel would be expelled, by choice of the velocity and certain other parameters to increase the freeboard temperature to as high as 2,500°F. This was not studied in any depth and was not tested.

(d) Combustion catalysts may properly be treated under temperature, since they increase the rate of the process just as an increase in temperature would.

Combustion Engineering, Inc. (CE) has a number of patents (18) in which combustion catalysts were used in a fluidized-bed to promote complete combustion. Pope, Evans and Robbins did not evaluate the methods of CE and in a number of discussions CE has not suggested application of their techniques.

It is likely that CE's methods are not compatible with coal firing.

A combustion catalyst has also been studied by Pope, Evans and Robbins and will be discussed in the report which will follow this one. Although the catalyst markedly increased the combustion efficiency, to the extent that the CBC was starved of fuel, it did not appear to make the Carbon-Burnup Cell unnecessary.

The conclusions drawn from this work and study was that it was not useful to attempt to increase combustion efficiency by operating the coal-burning fluidized-bed at a temperature higher than that required to provide a stable process and burn out CO and hydrocarbons.

### M.6.3 Turbulence

Turbulence implies any means for increasing the oxygen flux to the burning particles and includes high excess air as well as mixing techniques.

(a) The action of the fluidized-bed appears to provide sufficient turbulence to reduce the thickness of the boundary layer for heat transfer to tubes. It should also have a similar action on the boundary layer of carbon monoxide leaving a burning char particle. No means for increasing this action is apparent.

(b) A number of studies have shown that high combustion efficiencies could be achieved by operating at high levels of excess air but also at relatively low superficial velocities. The most recent to become available is that by Hoy (13).

Studies by Pope, Evans and Robbins indicate that even at high levels of excess air ( ≈50%), complete char burnout is not achieved in a low-temperature fluidized bed, operating at high velocities. Referring to Equation 42 an excess air requirement of 90 to 100% (RO = 12.7%) is indicated for a 99-1/2% combustion efficiency.

(c) Some early tests were made using over-fire air jets so as to provide turbulence in the freeboard. The results were not promising. Even if the tests had been successful, over-fire jets do not appear consistent with good fluidized-bed boiler design practice.

(d) Members of the staff of Combustion Engineering suggested that collected fines might be finely ground before recycle so as to increase the carbon surface area, the equivalent of increasing turbulence. This suggestion was never evaluated.

The conclusion drawn from this work and study was that it was not useful to attempt to increase mass transfer by increasing turbulence or increasing excess air.

## M.7 Invention of the Carbon-Burnup Cell

It appeared from all of these efforts that it was not feasible to provide the conditions required for single-pass char burnout and still retain enough of the potential advantages of a fluidized-bed boiler to make the development worthwhile. For reasons which are unknown, the obvious answer, the Carbon-Burnup Cell, was not obvious. Once it was accepted that the collected char had to be recycled, it made little sense to return it to the same bed from which it had originally come.

Possibly the fact that the FBM and FBC were next to one another in 1968 made the experiment self-suggesting.

Therefore, the water cooled FBC was internally insulated by a carbon-steel liner backed with high-temperature insulation. The FBM and FBC were then operated in tandem; the FBM at a low bed temperature and moderate excess air. The FBC, firing the FBM's collected fly ash, was operated at 1900°-2000°F with excess air being used to prevent exceeding 2000°F. A few ounces of coal were added each time the bed temperature fell below 1900°F. This test indicated that a combustion efficiency exceeding 99% could be achieved in a fluidized-bed boiler which contained a Carbon-Burnup Cell.

The Office of Air Programs asked that implications of the Carbon-Burnup Cell be more fully explored. This led to the work described in this report.